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Soda Ash Treatment of Neutralized Mine Drainage

Gwin, Dobson, and Foreman, Inc, Altoona, Pa

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16. ABSTRACT

Utilization of acid mine drainage (AMD) streams as a source of potable and industrial water has become a major goal of several proposed AMD treatment schemes. From among the various schemes available, the lime neutralization/soda ash softening process was selected for use at Altoona, Pennsylvania.

The treatment plant, as constructed, has the capability of treating waters from Kittanning Run (acid mine polluted) alone or in combination with waters from other city sources to achieve: (1) neutralization and iron removal to levels satisfactory for stream release, (2) softening to approximately 100 mg/l $CaCO_3$ hardness for municipal use, and (3) softening to a hardness of 200 mg/l $CaCO_3$ or higher to meet industrial use requirements.

The objective of this study was to evaluate the technical and econ mic feasibility of softening neutralized AMD waters by means of the cold lime/soda ash process. The study was conducted full-scale at the Altoona Treatment Plant located near the Horseshoe Curve area of Altoona, Pennsylvania. Unit processes employed at the plant consisted of lime neutralization, aeration, settling, soda ash softening, recarbonation, and filtration.

The results generally indicated that the desired quality could be achieved. Effluent quality of 100 mg/l hardness cost 10 cents/cu m (37 cents/1000 gal); treatment to 200 mg/l cost 9 cents/cu m (33 cents/1000 gal).

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FOREWORD

When energy and material resources are extracted, processed, converted, and used, the related pollutional impacts on our environment and even on our health often require that new and increasingly more efficient pollution control methods be used. The Industrial Environmental Research Laboratory - Cincinnati (IERL-Ci) assists in developing and demonstrating new and improved methodologies that will meet these needs both efficiently and economically.

This report is a product of the above efforts. It documents the technical effectiveness and cost of full-scale treatment of a mildly acidic mine drainage source by the lime/soda softening process to produce an effluent suitable for augmenting the domestic and industrial water supply for the City of Altoona, Pennsylvania. This was one of several projects undertaken by IERL-CI to develop and demonstrate acid mine drainage treatment and abatement processes to provide alternatives to government and industry in the selection and design of treatment facilities to meet the demands of the expanding extractive industry. Contact the Extraction Technology Branch for further information.

David G. Stephan
Director
Industrial Environmental Research Laboratory
Cincinnati

ABSTRACT

Utilization of acid mine drainage streams as a source of potable and industrial water has become a major goal of several proposed acid mine drainage treatment schemes. From among the various schemes available, the lime neutralization/soda ash softening process was selected for use at Altoona, Pennsylvania.

The treatment plant, as constructed, has the capability of treating waters from Kittanning Run (acid mine polluted) alone or in combination with waters from other city sources to achieve: (1) neutralization and iron removal to levels satisfactory for stream release, (2) softening to approximately 100 mg/l $CaCO_3$ hardness for municipal use, and (3) softening to a hardness of 200 mg/l $CaCO_3$ or higher to meet industrial use requirements.

The objective of this study was to evaluate the technical and economic feasibility of softening neutralized acid mine drainage waters by means of the cold lime/soda ash process. The study was conducted full-scale at the Altoona Treatment Plant located near the Horseshoe Curve area of Altoona, Pennsylvania. Unit processes employed at the plant consisted of lime neutralization, aeration, settling, soda ash softening, recarbonation, and filtration.

The study was conducted over a 5-month period from August to December 1974 and for a short time in 1975. The results generally indicated that the desired quality could be achieved. Effluent quality of 100 mg/l hardness cost 10 cents/cu m (37 cents/1000 gal); treatment to 200 mg/l hardness cost 9 cents/cu m (33 cents/1000 gal).

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The physical plant which was the subject of this study was constructed with "Operation Scarlift" funds, the $500 million conservation bond issue program administered by the Commonwealth of Pennsylvania through the Department of Environmental Resources. The actual study was performed for DER under grant agreement with the U.S. Environmental Protection Agency. Support of the project by these agencies is gratefully acknowledged.

Data collection, evaluations and the preparation of this report were under the direct supervision of technical staff of Gwin, Dobson & Foreman, Inc. and Mr. David A. Long, Associate Professor of Civil Engineering at Pennsylvania State University.

We wish also to acknowledge the splendid cooperation provided by the City of Altoona Treatment Plant Employees and their supervisor, Mr. David Barr in collecting samples, exercising the operational control, and, in general, providing the overall work effort essential to conducting a project of this type.

SECTION I

INTRODUCTION

Lime neutralization of coal mine drainage is a well known and widely applied process for upgrading stream water quality. When a significant portion of the watershed available to a municipality for water supply is affected by mine drainage pollutants, the question as to whether or not neutralized mine drainage can be economically treated to a more beneficial use level becomes a logical consideration. The $4.8 million water treatment plant at Altoona, located in south-central Pennsylvania, was constructed with "Operation Scarlift" funds* with a view to addressing both of these problems, i.e., stream pollution abatement and municipal water supply.

The watershed involved drains a portion of the mountainous area to the southwest of Altoona in the vicinity of Penn Central Railroad's well-known Horseshoe Curve. The after effects of deep mining coal in this region has rendered two streams acid, -- Glen White Run and Kittanning Run. The latter, with only half the drainage area, is about seven times more acid than Glen White Run. Early in Altoona's history, a series of three raw water reservoirs, together with transmission lines, were constructed to provide a water supply for the City during periods when the high quality source (Mill Run Reservoir) became depleted. Under normal conditions, the backup reservoir system in the Horseshoe Curve Area consisted primarily of the better source, Glen White, with Kittanning Run bypassing the impoundments altogether. During severe drought, however, there were times when the highest reservoir (Kittanning Point) was fed by the Kittanning Run source, and this poor quality of water was the only one available to City consumers.

The rationale for the new treatment plant was to provide two levels of treatment: 1) lime neutralization of Kittanning Run to meet stream release standards (pH 6.0 to 9.0 and iron 7.0 mg/l), and 2) lime neutralization and lime/soda ash softening of the reservoir system supply, fed by the Glen White source, to meet recommended (USPHS) potable water standards. Our study was prompted by the question, would it be technically and economically feasible to treat the Kittanning Run raw water source to a more beneficial-use quality? It was the main objective of this project to develop the basis for considering this question within the limits imposed by the available treatment units.

*Pennsylvania bond issue money ($500 million).

SECTION II

CONCLUSIONS

A basic requirement of the study method utilized was that a relatively constant hydraulic 13,200 cu m/day (3.5 mgd) and mineral loading should be applied to the treatment process. In this regard, certain qualifying factors should be mentioned prior to discussing the major findings of the study. First, since the Kittanning Run (K.R.) source is an open channel flow, weather (rainfall) caused some variation in raw water quality. This condition could be controlled using K.R. exclusively or by blending the total flow in K.R. with constant quality Impounding Dam (I.D.) water, a fairly successful procedure so long as K.R. flow did not exceed the maximum 13,200 cu m/day (3.5 mgd). Temperature, however, was a parameter which could not be controlled.

Another equally important factor was the inability to obtain consistent softener operation with reaction zone solids and good clarification. This situation was a two-edged problem inasmuch as it was necessary to evaluate nontypical process performance of the unit as well as delay filter startup because of this condition. Data collection of this process was thereby limited for fear of prematurely fouling the filter media. The following conclusions were drawn from the study:

1. Coal mine drainage (average characteristics of pH 3.0, acidity 160 mg/l, iron 14.5 mg/l, and manganese 4.5 mg/l) was treated by the neutralization and lime/soda ash process to produce a finished water quality that will generally meet the EPA National Interim Primary Drinking Water Regulations.

2. The cost of producing a water suitable for blending with a potable supply from the subject drainage source is about 10¢/cu m (37¢/1000 gallons). This cost includes power, chemicals, operating personnel and maintenance.

3. Lime addition to the neutralization process did not contribute an equivalent amount of calcium hardness in the sedimentation basin effluent. Some calcium (approximately 15 mg/l) was apparently precipitated as calcium sulfate, as further evidenced by an equivalent sulfate loss, in the neutralization stage.

4. Neutralization of the raw water source resulted in a clarified effluent averaging an 85-percent reduction in total iron, i.e., 14.5 mg/l to 2.2 mg/l. The major portion of the remaining iron

was present as filterable suspended solids with only 0.1 mg/l dissolved iron.

5. The light fragile iron hydroxide floc formed in the slow mix and aeration tanks was easily broken up by turbulent flow caused by the fall from the aeration tank effluent weir into the collection trough. Settling and iron removal were adversely affected by this condition. Adding coagulant aid (cationic polymer) to the sedimentation basin influent significantly improved settling and resulted in a more compact sludge.

6. The most effective removal of manganese was accomplished by the addition of potassium permanganate to the raw water influent line ahead of the neutralization stage. If no potassium permanganate was used, the most effective manganese removal was observed at pH 11.0.

7. Magnesium removal (raw water magnesium 18 mg/l) in the neutralization stage was pH dependent according to the following schedule: pH 7.0 resulted in very little reduction, pH 9.0 to 9.5 caused a reduction of about 50%, and pH 11.0 to 11.5 resulted in removal of practically all of the magnesium (1 mg/l remaining).

8. The total aluminum applied to the softening stage was fairly constant, averaging 2.5 mg/l. Soluble aluminum, however, varied with pH in the sedimentation basin effluent from 1 mg/l at pH 7.0 to 2.5 mg/l at pH 11.0.

9. The softener unit operation employed was of the solids contact type that requires the return of settled solids to the reaction zone for intimate mixing of previously formed floc with the applied flow and chemical addition. The buildup of reaction zone solids (10-15% settleable solids by volume in 15 minutes) was a prerequisite for efficient reaction and effective clarification according to the manufacturer. With the exception of approximately 3 weeks during the study, it was impossible to obtain the recommended type of softener operation. The reason for failure of the unit to buildup solids is not understood.

10. The operating conditions under which reaction zone solids could be generated required a pH of 11.0 or higher. During these test periods the neutralization stage pH was 7.0 to 8.0. Heavy floc carryover from the sedimentation basin was characteristic of this first-stage operation. It was determined, however, that good softener operation at pH 11.3 also could be obtained without the presence of heavy floc carryover from the first stage. The favorable softener operation with a desired level of reaction zone solids was accomplished during periods when the water temperature was above 12°C. At water temperatures below 10°C, the desired softener process operation could not be reproduced.

11. The general appearance of the softener effluent for the most part could be characterized as a turbid, homogeneous quality of water containing a finely divided precipitate. Samples drawn from various levels and zones within the tank failed to show any distinct differentiation in the liquid contents which would indicate significant solids buildup.

12. The use of alum and/or coagulant aids (CA 233 and CA 253) was not effective in improving the coagulation - flocculation process in the softener.

13. The capacity of the recarbonation unit was not sufficient to lower the pH of the softener effluent below 9.5 when the softener process was operated at a pH in excess of 11.0. The carbon dioxide feed capacity was 682 kg/day (1500 lb/day) or about 51 g/cu m (428 lb/MG) of water to be treated.

14. A severe filter plugging problem was experienced during the period from October 23 to 28. It was suspected that the cause for this condition was due, at least in part, to coagulant aid fouling of the top 10 to 15 cm (4 to 6 in) of the filter media.

15. Settled solids in the thickener could not be effectively moved to the center collection well. The fluid characteristics of the sludge resulted in a condition whereby the solids accumulated in a ring approximately one third to half way from the periphery of the tank. Air lancing was partially successful in moving the sludge to the tank-center. Ultimately, it was necessary to lower the level in the thickener and hose the solids toward the middle sump.

16. The total amount of metered lime fed to the process could be accounted for based on theoretical lime requirements and observed results.

SECTION III

RECOMMENDATIONS

It should be clearly understood that the treatment units and the process discussed in this report were not designed or intended to treat the Kittanning Run source of mine drainage to a degree better than that required for stream release, i.e., pH 6.0 - 9.0 and iron 7.0 mg/l or less. However, having the opportunity to observe the complete system with Kittanning Run as the raw water source has provided valuable operating experience relative to applying better treatment if the need arises. It is in reference to this possible application, as well as other general problem areas, that the following recommendations are made:

1. Conventional design criteria will generally apply to the softening process (lime/soda ash) so long as the water temperature remains above 54°F (12°C). It is assumed that if Kittanning Run or a similar quality of raw water were to be treated, impoundment would be provided, so in all probability, temperature below this threshold level would not be experienced.

2. Turbulence should be avoided at transfer points in the system after mixing and particularly ahead of the clarifier to minimize floc breakup.

3. The use of potassium permanganate should be limited to head-end of the plant application. This study indicated that the addition point ahead of neutralization (7 hours before the softening unit) was best. Potassium permanganate feed directly to the softener was very difficult, if not impossible, to control.

4. Application of the particular solids contact type of softener unit used in this study appears questionable.

5. Design of the thickener unit operation for handling a mixture of iron hydroxide/calcium carbonate solids should take into consideration the tendency of this type of sludge to "doughnut," or resist movement to the tank center. The bottom slope in the unit available for the study was about 14.5%. Increasing the floor slope would facilitate sludge removal and solids control.

6. Provision should be made for feeding coagulant aid to the sedimentation basin (ahead of softening) inlet line. This capability is important because of the characteristic light iron hydroxide floc

formed in the neutralization stage. In general, flexibility in chemical addition points is most desirable.

7. Our data analysis indicates that though on occasion a softener effluent hardness of less than 100 mg/l $CaCO_3$ was obtained, we could not expect an average of much less than 120 mg/l, regardless of the amount of soda ash fed. This result must be interpreted with some caution because the process condition did not afford control of temperature, softener reaction zone solids, or continuous filter operation. Where a specific application of this treatment process is being considered, it would be advisable to determine this minimum achievable hardness level in the laboratory.

SECTION IV

PROCESS DESCRIPTION

The treatment system available for study under this project is shown diagrammatically in Figure 1. The operational units employed are grouped together according to the major process objectives, neutralization, softening and sludge handling. The first stage, or neutralization process, consists of the flash mix, slow mix, aeration and sedimentation units. Detention times in this flow-through sequence are 25 seconds, 25 min, 1.1 hr, and 6.17 hr respectively based on a flow of 13,200 cu m/day (3.5 mgd). Settled solids are pumped from the sedimentation basin at a controlled rate to the flash mix (recycled sludge) and to the thickener clarifier (waste solids). Lime slurry is fed to the flash mix in a controlled amount to maintain the pH called for in the slow mix tank. In addition, as a result of this investigation, provision has been made to feed potassium permanganate into the influent raw water line.

Effluent from the first stage sedimentation basin meets stream release water quality and, in the case of Kittanning Run, would normally be discharged from the plant. For purposes of this study, however, the effluent was directed to the mix box where lime can be added prior to softening and recarbonation. The softener unit is a circular tank arranged with separate compartments to provide zones for rapid mix, flocculation, and clarification (Walker Process Clariflow Unit). In theory, the unit is designed to pump settled solids from the bottom-center of the tank back to the reaction zone. Chemicals (soda ash, alum, coagulant aid, and potassium permanganate) can also be fed at this point to allow for mixing of the applied flow with previously formed floc and chemicals so that the required reactions, coagulation, and flocculation can occur. Theoretical residence times for the reactor and settling zones are 25 minutes and 83 minutes respectively. Flow from the softener is to the recarbonation chamber 12.2m x 2.4m x 3.0m (40'x 8'x 10') where carbon dioxide is added. Recarbonation detention time for the 13,200 cu m/day (3.5 mgd) flow is 10 minutes. The effluent from the recarbonation unit is filtered prior to clearwell storage and chlorination.

The major components of the sludge handling process are the thickener clarifier, holding tank and vacuum filters. For purposes of this discussion the backwash lagoons, which receive backwash water from the filters, filtrate from the vacuum filter, and thickener overflow, also are included in the solids handling operation. The thickener receives underflow solids from the first stage (sedimentation basin) and the second stage (softener). Thickened sludge is pumped to a holding tank and from there to the vacuum

filters as required. The vacuum filters are elevated so that the sludge cake can be conveyed to a hopper and periodically loaded on a truck for hauling to final disposal.

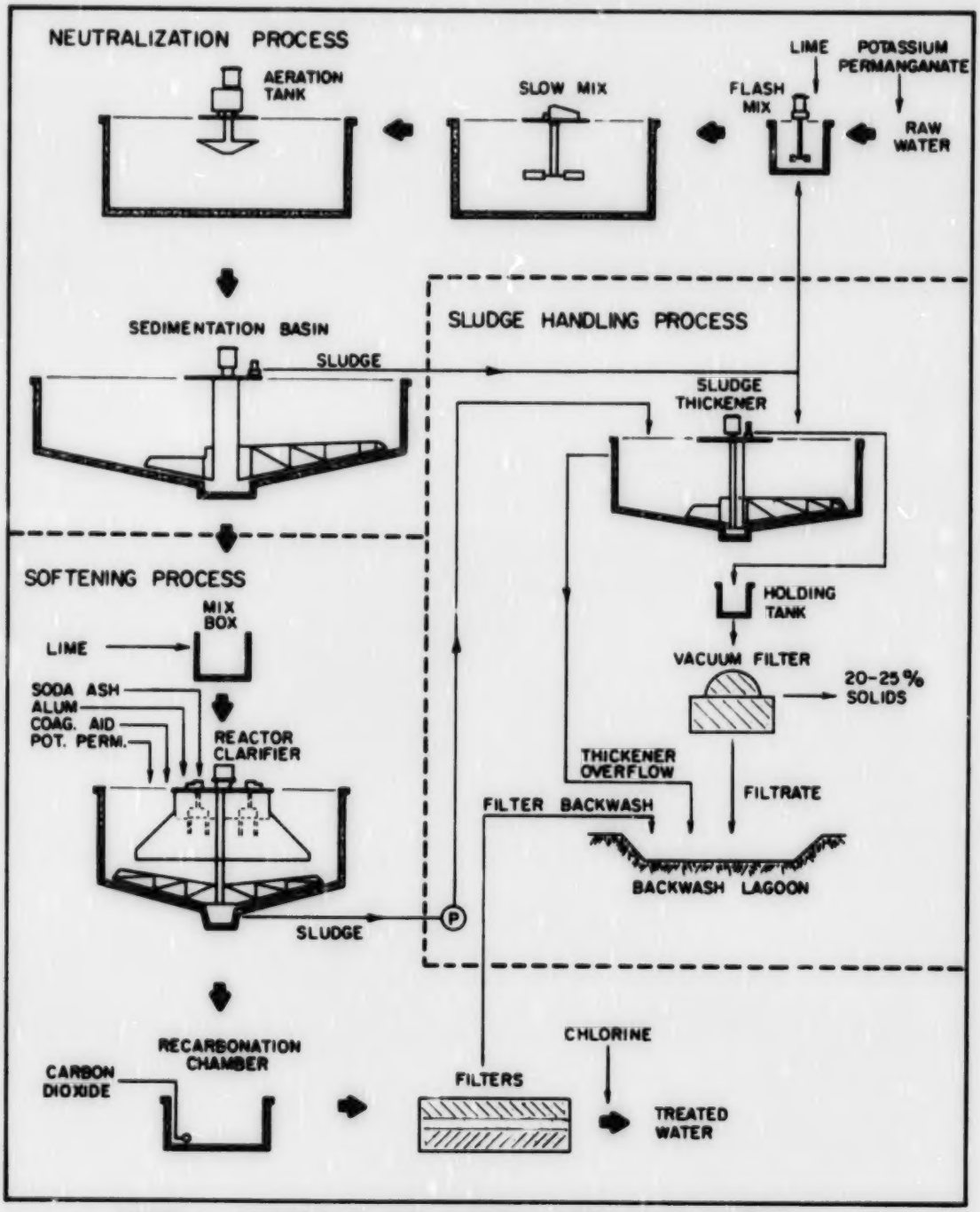

FIGURE I. DIAGRAM OF THE ALTOONA TREATMENT SYSTEM ILLUSTRATING UNIT PROCESSES

SECTION V

PROCESS ENGINEERING AND OPERATION

PROJECT PLAN

The project was divided into major phases. The first phase was designed to achieve an effluent quality which would meet the requirements of the 1962 Public Health Service Drinking Water Standards (1), except for sulfate content. Specific objectives were to produce a finished water with approximately 100 mg/l $CaCO_3$ hardness and containing no more than the permissible 0.3 mg/l and .05 mg/l of iron and manganese respectively. This product water was designated as "municipal water supply quality" (MWSQ). The 1962 Standards were selected as the basis for process control because the interim regulations being developed under the "Safe Drinking Water Act" (2) by the Environmental Protection Agency do not include "secondary standards" for those parameters of concern in this study.

The second major phase of the project was intended to study the production of an effluent designated as "industrial water use quality" (IWUQ). The effluent target hardness for this phase was 200 mg/l $CaCO_3$. Iron and manganese levels were to be the same as for the municipal supply quality phase.

In addition to the two major phases described, project data were to be analyzed to determine the ability of the first stage process, i.e., aeration and neutralization, to produce an effluent that would be suitable for stream release under the Commonwealth of Pennsylvania, Department of Environmental Resources regulations (3). In order to meet these requirements (pH 6.0 to 9.0, Fe less than 7.0 mg/l, and alkalinity greater than acidity), the process would be operated somewhat differently than was done for the two major phases described above.

It was possible, however, to analyze the data collected and to predict the ability of the first-stage process to meet the requirements indicated.

Finally, although not a major consideration in this study, some information on sludge production was generated and will be evaluated briefly in a later section.

During each phase of the study, operational parameters were varied in an attempt to determine the optimum operating conditions to produce the desired effluent quality for that phase of the study. The primary parameters which were varied were pH, amount of chemicals added, and the point of chemical application. Various pH's were set at different stages in the over-

all treatment process to see which scheme would minimize operating problems while producing an acceptable effluent quality. To clarify this further, during parts of the study the desired process pH was achieved initially in the neutralization stage and then was carried right on through the softening stage while in other parts the pH's were varied in the first stage unit as a pre-treatment (iron and manganese oxidation stage) and the pH's for the softening reactions were regulated in the reactor-clarifier unit.

While it would have been highly desirable to have been able to obtain filter operating data for the duration of the project, it was not possible to do so under the operating conditions encountered. Therefore, the majority of the data analysis presented will be in terms of the softener and recarbonation basin effluent characteristics only. Filtration of the recarbonation basin effluent would undoubtably yield effluents with somewhat different charactertistics than those reported herein but it was impossible to gather such data during this study other than for a limited period during the second phase. Generally, filtration would be expected to produce water with lower hardness, iron,and manganese than reported due to "seasoning" of the filters.

The specific goals of the study were as follows:

1. Determination of the chemical dosages required to yield a finished water which will meet specific water use criteria including:

   (a) stream quality standards
   (b) municipal water supply meeting the USPHS recommended limits (100 mg/l hardness), and
   (c) industrial usage water use quality, 200 mg/l hardness.

2. Evaluate operational problems associated with the process.

3. Determine the operating costs relevant to obtaining various qualities of product water.

4. Evaluate the feasibility and economics of utilizing the lime/soda ash treatment process for municipal and industrial uses on acid mine drainage.

RAW WATER QUALITY

Raw water for the study was a blend of Kittanning Run (K.R.), a moderate acid mine drainage affected source, and Impounding Dam (I.D.), a better quality water. Design flow through the portion of the plant used for the study was 13,200 cu m/day (3.5 mgd). Because the sustained yield of Kittanning Run is less than 13,200 cu m/day (3.5 mgd), it was necessary to blend the two sources as indicated in order to achieve the design flow. Table 1 shows the blended water characteristics for a blend typical of that treated during the study, i.e , 5680 cu m/day (1.5 mgd) of K.R. plus 7570 cu m/day (2.0 mgd) of I.D.

TABLE 1. TYPICAL BLENDED RAW WATER CHARACTERISTICS
KITTANNING RUN : IMPOUNDING DAM = 1 : 1.33

Parameter	Value
pH	3.0
Acidity, mg/l as $CaCO_3$	170
Calcium, mg/l	28
Magnesium, mg/l	18
Iron, mg/l	17
Manganese, mg/l	4.5
Aluminum, mg/l	13
Sodium, mg/l	1.8
Hardness, mg/l as $CaCO_3$*	260
Sulfates, mg/l	270
Specific conductance, umhos/cm^2	820

*Calculated hardness value.

ANALYTICAL METHODS

Samples for the various analyses performed during the study were collected daily and composited by treatment plant personnel. Following collection, the samples were transported to the laboratory for analysis.

EPA (4) methods were used for the determination of acidity, alkalinity, suspended and dissolved solids, aluminum, calcium, iron, magnesium, manganese, sodium, and sulfate.

Standard Methods (5) were used to determine pH and specific conductance. All hardness values reported herein were calculated in accordance with Standard Methods using the values obtained for Ca, Mg, Al, Mn, and Fe.

DATA PRESENTATION

The nature of this study was such that a vast amount of data was collected under varying operating conditions. Also, because the study was conducted at full plant scale, process control was more erratic than would be expected if conducted at laboratory or pilot scale. The data, therefore, showed considerable variability. All operating data obtained during the course of the study were included in the data analysis unless there was an

obvious error in sampling or laboratory analysis. It was felt the data could be presented best in the text of the report by means of probability plots that show the statistical correlation of the data. Summaries of the complete operating data are presented in Appendix B.

Data analysis for the preparation of the probability plots was done in accordance with the procedures outlined in the Handbook for Monitoring Industrial Wastewater (6). Data which are normally distributed yield a straight line when plotted on probability paper. In some instances, abnormal breaks in the data are observed on such plots. This observation indicates that some change in operating conditions occurred which resulted in data being obtained from a different distribution and hence, they do not plot along the original straight line. These data may, however, plot on a different straight line showing a normal distribution of data obtained under the changed operating conditions.

STREAM RELEASE QUALITY RESULTS

As previously indicated, the requirements for stream discharge of acid impregnated water are: (1) pH between 6.0 and 9.0 and (2) dissolved iron not to exceed 7.0 mg/l. Only the first stage of the plant (neutralization process, Figure 1) would be required to achieve the necessary degree of treatment for stream release. Because the plant was not operated solely to produce stream release quality effluent during the study, however, the neutralization stage was operated as a pretreatment step to the softening operation. As will be discussed in more detail in later sections, the pH in the neutralization stage was varied from 7.0 to 11.0 to investigate the effect of this parameter on the subsequent softening operation. The data obtained during the study, therefore, do not reflect operating conditions comparable to those which would be employed for acid mine drainage treatment only.

Figures 2 and 3 are probability plots of the raw water pH and the sedimentation basin effluent pH from the neutralization stage respectively. Figures 4 and 5 are similar plots for total iron. As can be seen from Figure 3, the stream release quality goal of a minimum pH of 6.0 was met 95 percent of the time. The upper pH limit of 9.0 would not be of concern if the treatment plant was operated only to meet stream release quality criteria. The high pH's used in this study were necessary for the softening processes used to meet the other water quality goals. Accurate control of pH in the range of 7.0 to 9.5 was impossible during the study so all data for target pH's in this range were treated together for use in preparing Figure 3. Control was much better for a target pH of 11.0 as shown by the lesser scatter for these data in Figure 3. No definite reason for the break in the curve for the raw water iron concentration as shown in Figure 4 was apparent, although it may have been related to flow fluctuations with resulting dilution and concentration of iron in Kittanning Run. Figure 5 shows the stream release quality goal of 7.0 mg/l of iron was met 100 percent of the time.

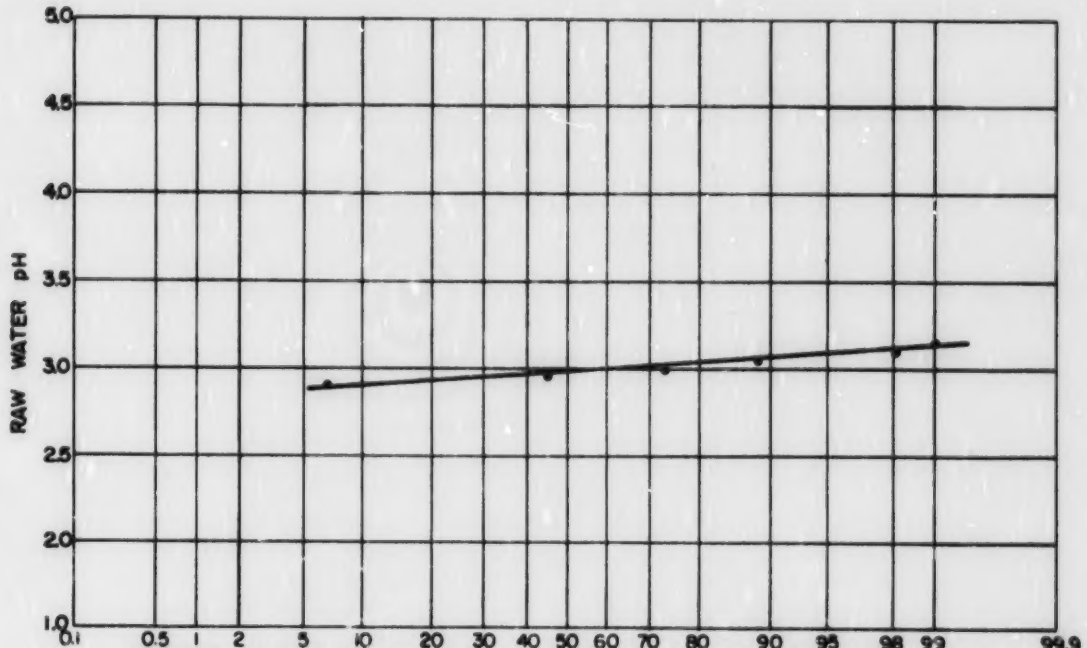

FIGURE 2: RAW WATER pH PROBABILITY PLOT

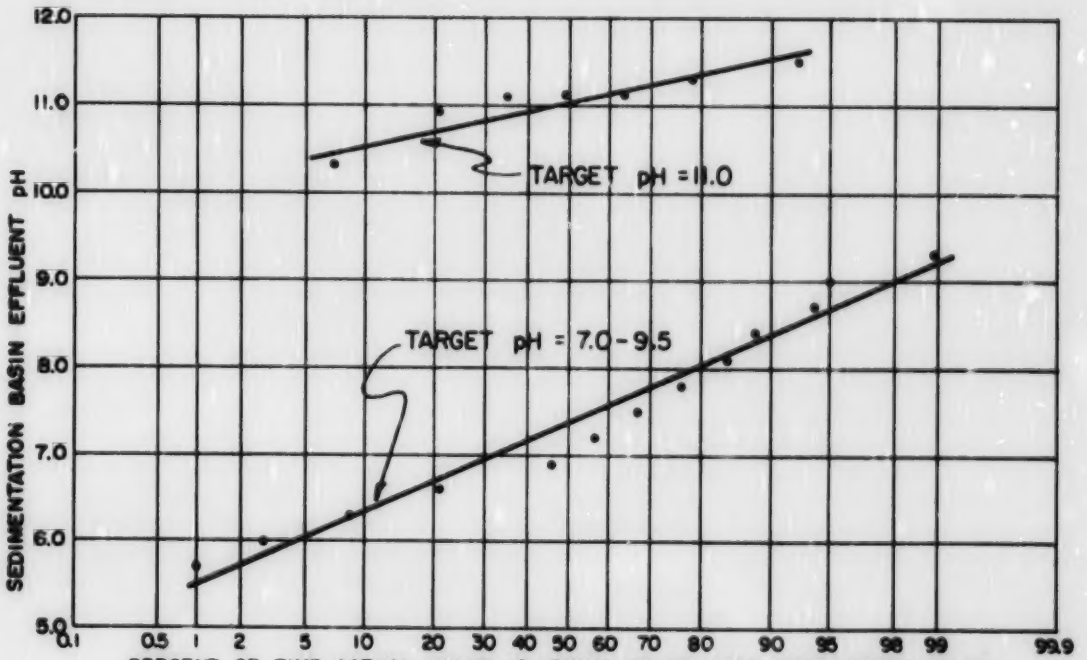
SEDIMENTATION BASIN EFFLUENT pH
12.0
11.0
10.0
9.0
8.0
7.0
6.0
5.0
TARGET pH = 11.0
TARGET pH = 7.0 - 9.5
0.1 0.5 1 2 5 10 20 30 40 50 60 70 80 90 95 98 99 99.9
PERCENT OF TIME ACTUAL VALUE IS EQUAL TO OR LESS THAN GRAPH VALUE

FIGURE 3: SED. BASIN EFFLUENT pH PROBABILITY PLOT

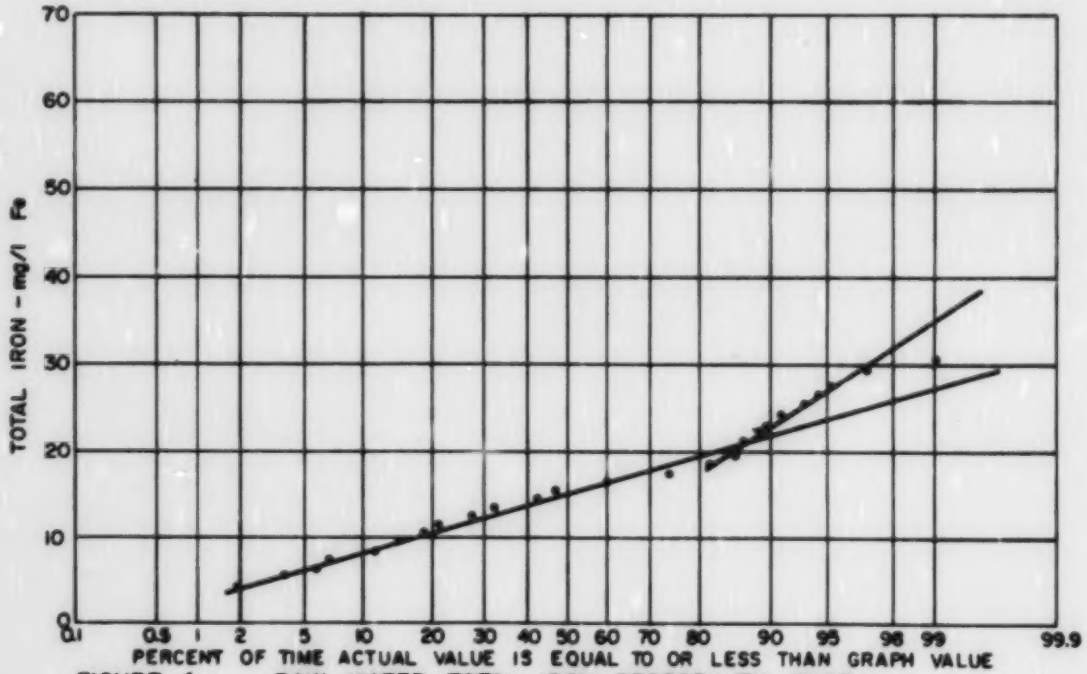

FIGURE 4: RAW WATER TOTAL IRON PROBABILITY PLOT

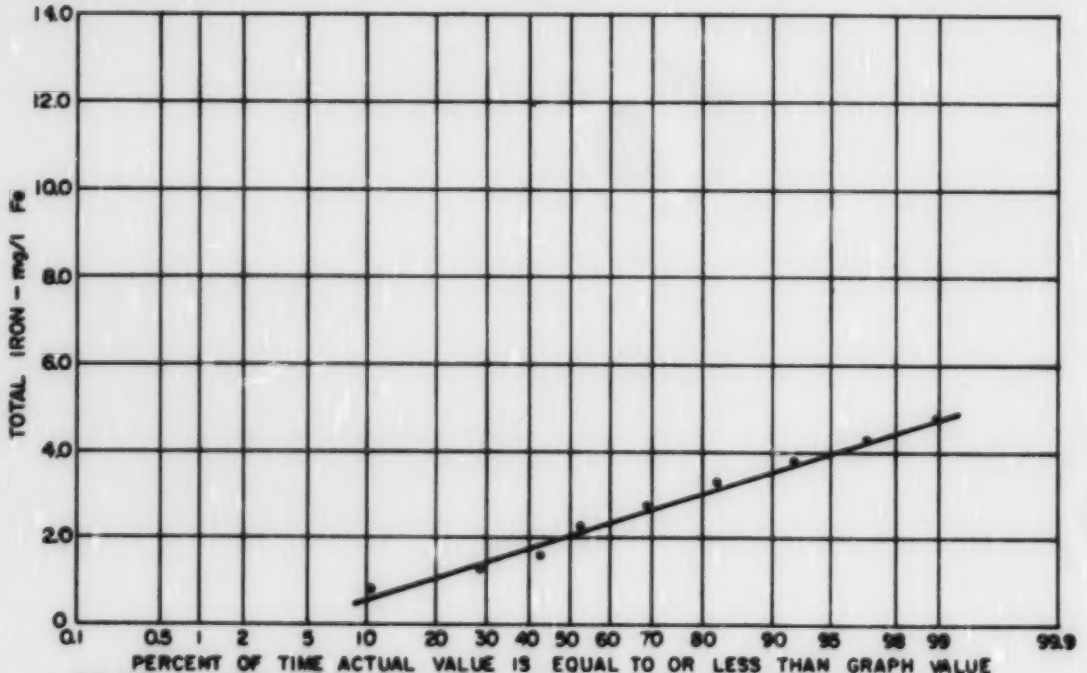

FIGURE 5: SED. BASIN EFFLUENT TOTAL IRON PROBABILITY PLOT

Optimization of the neutralization stage to meet stream release requirements would have to be done independently of the softening operation in order to develop the most cost effective mode of operation. It was observed that when the pH in the neutralization stage was maintained at approximately 7.0 there was some carryover of large floc particles from the first stage sedimentation basin. An attempt was made to reduce this floc carryover in order to determine its effect on the operation. Any such reduction in floc carryover should result in lower total iron residuals in the effluent from this stage and hence would be advantageous in meeting the stream discharge requirements. An attempt was made to improve solids separation in the first stage by partially closing the valve in the sedimentation basin inlet line. This change was made to maintain a higher water level in this effluent channel from the aeration basin. Floc breakup should have been reduced under these operating conditions because free fall into the effluent weir trough was eliminated. This modification in operating conditions did not reduce the floc carryover significantly. In a further attempt to improve clarification, a coagulant aid (Calgon 233) was added into the sedimentation basin influent line while operating under the two conditions described above. The dosage of CA 233 was set at 1 mg/l, the maximum concentration permitted by EPA (7). Some improvement in clarification was noted but not enough data were collected to make a comparative analysis of its effect on total iron removal.

MUNICIPAL WATER SUPPLY QUALITY RESULTS

For the purposes of this project, municipal water supply quality was defined as treated water having a total dissolved solids (TDS) content of less than 500 mg/l. Further, the target hardness for the treated water during this portion of the study was 100 mg $CaCO_3$/l. Also, as indicated earlier, effluent limits for iron and manganese were 0.3 mg/l and 0.05 mg/l, respectively.

During this phase of the study, data were collected over a total of 74 days which were divided into 11 different operating periods. Process control conditions for these periods are summarized in Table 2. Because of operating problems with the softening unit, which will be discussed in more detail later, it was deemed advisable to vary pH's and other operating parameters as shown in an attempt to improve performance of the softening unit.

TABLE 2. PROCESS CONTROL CONDITIONS FOR MUNICIPAL WATER SUPPLY QUALITY PHASE

Period	Target pH		
	Neut. Stage	Soft Stage	
Aug. 12-19, 1974	11.0	11.0	Recarbonation; no filtration; tried Calgon 233; No measurable reaction zone solids.
Sept. 1-15	9.5	11.0	Recarbonation; no filtration; alum addition to Softener; no measurable reaction zone solids.
Sept. 16-19	7.0	10.5	Recarbonation; no filtration; no measurable reaction zone solids.
Sept. 20-29	7.0	11.4	Recarbonation not effective; no filtration; reaction zone solids started to buildup.
Oct. 1-4	7.0	11.4	No recarbonation or filtration, modified operation of aeration tank to investigate effect of floc carry-over. Reaction zone solids 14% by volume.
Oct. 5-7	7.0	11.5	Same operation as Oct. 1-4 except added Calgon 233; Reaction zone solids 13%.
Oct. 8-10	7.0	11.3	No recarbonation or filtration, reduced aerator operation; Reaction zone solids 12%
Oct. 12-16	9.5	11.3	Same operation as Oct. 8-10. No measurable reaction zone solids. Tried Calgon 253 on 11/16.
Nov. 25-30	8.5	10.5	Recarbonation (2 days); filtration.
Dec. 3-8	7.0	11.3	Recarbonation (4 days); filtration (2 days).
June 16-26, 1975	9.0	9.6	Recarbonation; no filtration. No measurable reaction zone solids

Examination of the characteristics of the typical blended raw water in Table 1 indicates that about 28 percent of the total hardness (TH) is due to magnesium (73 of 255 mg/l as $CaCO_3$). The target hardness of 100 mg $CaCO_3$/l, therefore, should have been achievable by leaving most of the magnesium in the water and by removing calcium hardness down to its minimum practical solubility of 30 to 40 mg $CaCO_3$/l. One concern, however, with this approach is the potential problem of magnesium silicate scaling in the high temperature service, such as boiler feedwater application, when the magnesium hardness exceeds 40 mg $CaCO_3$/l (8). Assuming magnesium hardness of approximately 70 mg $CaCO_3$/l is acceptable, the desired hardness should have been achieved by maintaining the softener pH at approximately 9.5 and by using conventional lime/soda ash softening operational parameters. Soda ash addition to remove non-carbonate hardness (NCH) could then be used to control the effluent total hardness.

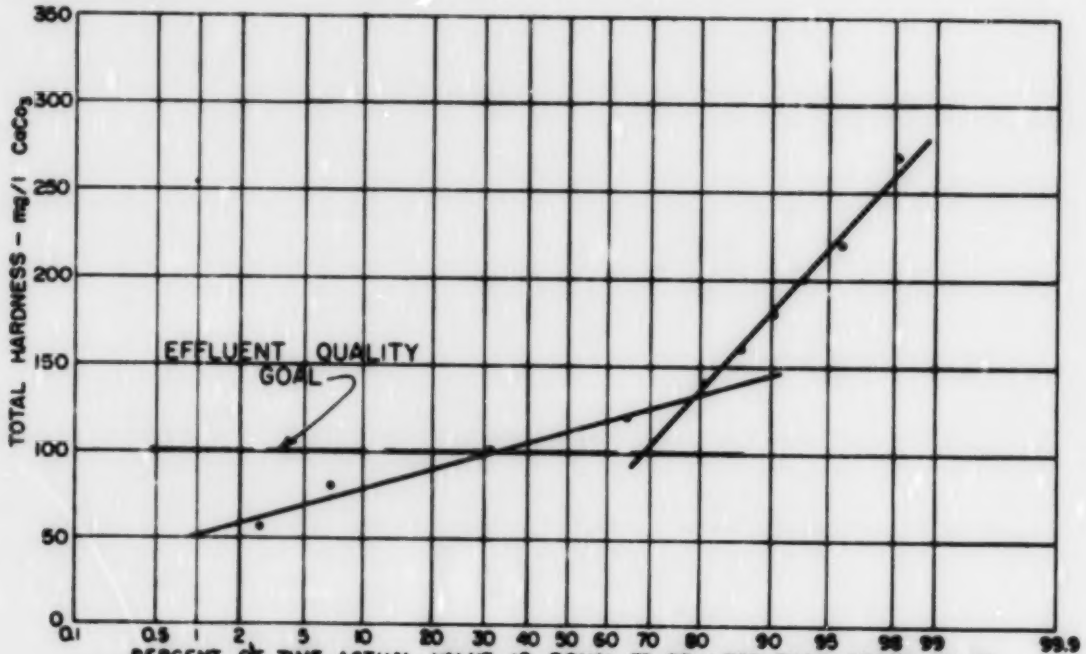

FIGURE 6: SOFTENER EFFLUENT TOTAL HARDNESS MUNICIPAL WATER SUPPLY QUALITY PROBABILITY PLOT

Figure 6 is a probability plot of the softener effluent total hardness for all data collected during this phase of the study. As indicated by this figure, the target hardness of 100 mg $CaCO_3$/l was achieved only approximately 30 percent of the time. Effluent hardnesses of 125 mg $CaCO_3$/l or less were obtained approximately 70 percent of the time and 85 percent of the time the effluent hardness was 150 mg $CaCO_3$/l or less. It is believed the primary reason for the poor performance of the softening operation was the inability to adequately build up and retain solids in the reaction zone of the Clariflow unit. Successful solids-contact softener operation is highly dependent upon having enough preformed crystals in the reaction zone to increase the rate of chemical reactions in the process. This point will be discussed further in a later section. The problem of maintaining pH control in the

range of 7.0 to 9.5 discussed previously made proper control of the softening process difficult and this lack of control may be part of the reason for poor overall performance. Because of the inability of the recarbonation unit to reduce the pH to desired levels during some operating conditions it was not possible to fully evaluate the effect of recarbonation on effluent hardness. As was discussed earlier, it was not possible to operate the filters during this portion of the study so the effect of filtration of the softener effluent was not obtained.

Figure 7 is a probability plot of the softener effluent total iron data collected during this phase of the study. This plot shows that the drinking water standard of 0.5 mg/l was achieved only 70 percent of the time. All dissolved iron values, however, were less than 0.2 mg/l and a total of 61 out of 72 data points for dissolved iron showed less than 0.1 mg/l of iron so effective filter operation should yield an effluent that would meet the standards without any difficulty.

Softener effluent total and dissolved manganese probability plots are shown in Figure 8. As can be seen from these plots, the drinking water standard limit of 0.05 mg/l was achieved 65 or 80 percent of the time depending on whether total or dissolved values were used. Manganese oxidation, which is normally the rate limiting step in the removal of manganese from potable water, proceeds rather slowly at pH's between 8.5 to 9.5. During those periods of the study when the neutralization stage was being operated at pH's less than 8.5 and because of the poor pH control achieved, effluent manganese values were often too high to meet the standards. It is entirely possible that filter effluent manganese values would be significantly lower than those observed due to the catalyzing effect of "seasoned" filters on manganese removal. As stated previously, however, it was not possible to evaluate this during the study.

Data were not collected routinely for those substances listed in the national interim primary drinking water regulations developed by EPA (9), but the results of one set of analyses (reported in Table 3) indicate that effluent from the plant could be expected to meet the standards for those substances that were measured. Not all of the substances included in the proposed standards were included in the analyses performed.

INDUSTRIAL WATER USE QUALITY RESULTS

As was indicated earlier, for the purposes of this study, industrial use quality water was defined as treated water with total dissolved solids in excess of 500 mg/l and a hardness of approximately 200 mg $CaCO_3$/l. Effluent limits for iron and manganese were the same as for the municipal water supply quality phase. This phase of the study was divided into six different operating periods which included a total of 34 days of operation and data collection. The process control conditions for these periods are summarized in Table 4. As noted in the table, the major changes in operation compared with that used for the first phase of the study were: 1) the addition of potassium permanganate to facilitate oxidation of manganese and 2) the operation of the filters for the major portion of this phase. Of course, the soda ash dosage was reduced from that used in Phase I in order to achieve

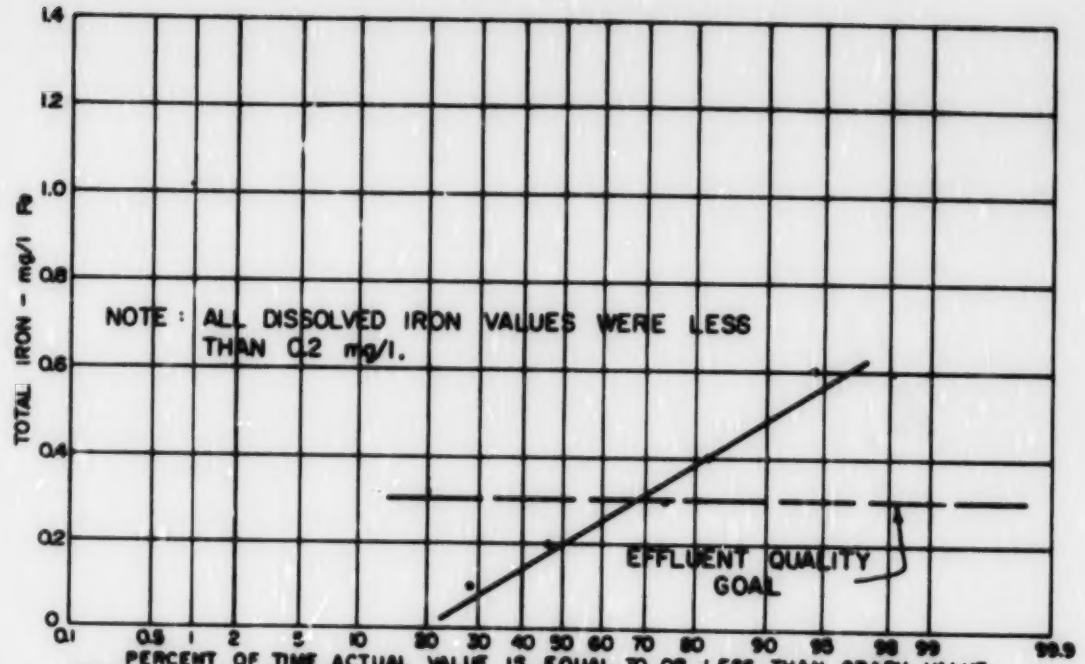
TOTAL IRON - mg/l Fe
1.4
1.2
1.0
0.8
0.6
0.4
0.2
0
NOTE : ALL DISSOLVED IRON VALUES WERE LESS THAN 0.2 mg/l.
EFFLUENT QUALITY GOAL
0.1 0.5 1 2 5 10 20 30 40 50 60 70 80 90 95 98 99 99.9
PERCENT OF TIME ACTUAL VALUE IS EQUAL TO OR LESS THAN GRAPH VALUE

FIGURE 7 : SOFTENER EFFLUENT TOTAL IRON MUNICIPAL WATER SUPPLY QUALITY PROBABILITY PLOT

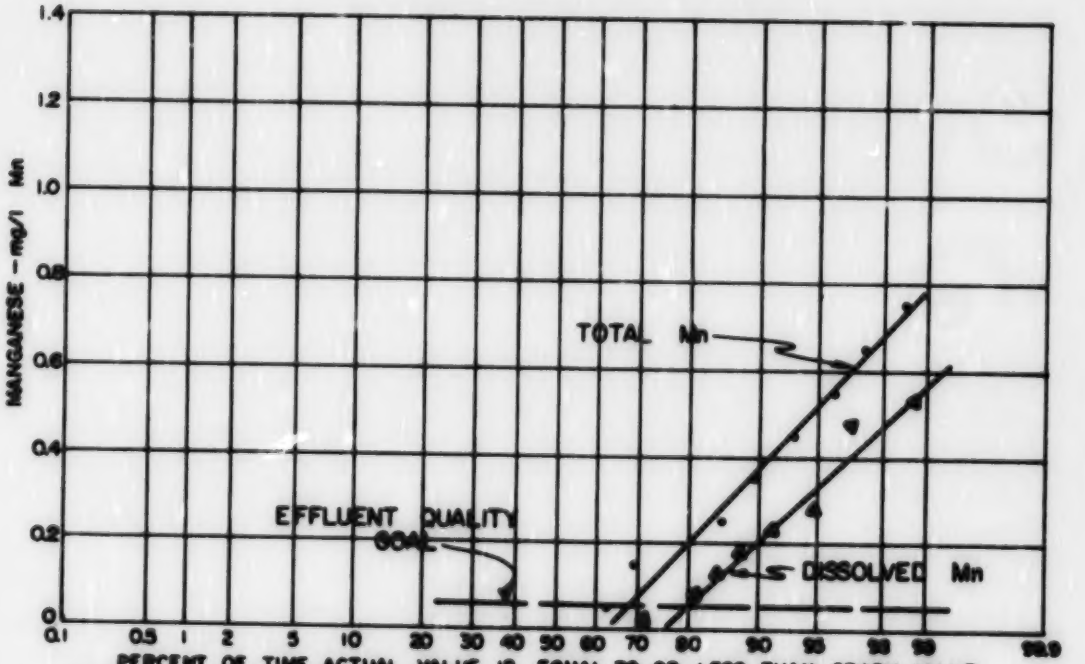

FIGURE 8: SOFTENER EFFLUENT MANGANESE MUNICIPAL WATER SUPPLY QUALITY PROBABILITY PLOT

the desired high effluent hardness.

TABLE 3. WATER QUALITY DATA COMPARISON WITH EPA NATIONAL INTERIM PRIMARY DRINKING WATER REGULATIONS

Parameter(1)	Raw water		Neutralization stage	Recarb. basin eff.	Proposed (8) int. Primary standards
	Kitt. Run	Imp. Dam			
Arsenic	0.030	0.030	0.030	0.030	0.05
Barium	0.05	0.05	0.05	0.05	1.
Cadmium	0.004	0.004	0.004	0.004	0.010
Chromium	0.010	0.010	0.010	0.010	0.05
Cyanide	0.005	0.005	0.005	0.005	0.2
Lead	0.030	0.030	0.030	0.030	0.05
Mercury	--	--	--	--	0.002
Nitrate	--	--	--	--	10.0
Selenium	0.010	0.010	0.010	0.010	0.01
Silver	--	--	--	--	0.05
Fluoride	0.33	0.19	0.29	0.29	1.4 - 2.4
Copper	0.3	0.020	0.010	0.010	1.0*
Zinc	1.1	0.15	0.070	0.030	5.0*

NOTE: All values are mg/l; the test results are from Reference 10.

* 1962 USPHS Drinking Water Standards.

TABLE 4. PROCESS CONTROL CONDITIONS FOR INDUSTRIAL WATER USE QUALITY PHASE

Period	Target pH		Comments
	Neut. Stage	Soft Stage	
Oct. 17-22	9.5	11.4	No recarbonation, filtration. Calgon 253 added @ 1 mg/l
Oct. 28 - Oct. 31	8.5-9.0	11.3	Recarbonation; no filtration
Nov. 1-10	8-5-9.0	11.3	Recarbonation and filtration
Nov. 11-17	8.5	11.3	Recarbonation (4 days), filtration; $KMnO_4$ feed (10 mg/l) into I.D. Influent line.
Nov. 18-21	8.5	10.5	No recarbonation, filtration, $KMnO_4$ feed (10 mg/l) into I.D. influent line
Nov. 22-24	8.5	10.5	Recarbonation (2 days) filtration

Figure 9 is a probability plot of the softener effluent total hardness for all of the data collected during this phase of the study. As can be seen from this figure, the target hardness of 200 mg $CaCO_3$/l was achieved approximately 65 percent of the time with an effluent hardness of 225 mg $CaCO_3$/l or less approximately 90 percent of the time. Comparison of filter effluent (clearwell) data with the softener effluent data for that portion of Phase II when the filters were in operation showed approximately 4.5 percent reduction in hardness through the filters (192 mg $CaCO_3$/l in the clearwell compared with 201 mg $CaCO_3$/l in the softener effluent based on weighted mean values).

Iron and manganese data obtained for this phase of the study were quite similar to those obtained during the municipal water supply quality phase and presented earlier. All values for dissolved iron in the softener effluent were below 0.1 mg/l. The weighted mean value for total iron in the softener effluent during the period the filters were in operation was 0.55 mg/l as compared to 0.18 mg/l in the filter effluent. These values show that filtration resulted in a 67.3% reduction in total iron and resulted in levels well below the drinking water goal of 0.3 mg/l. Only two values for dissolved manganese exceeded 0.05 mg/l during this period of data collection. Filtration of the softener effluent resulted in appreciably lower values for total manganese as evidenced by the weighted mean value of 0.03 mg/l for filter effluent as compared with 0.19 mg/l for the softener effluent; an 84.2 percent reduction.

Figures 10 and 11 are probability plots of the total dissolved solids and sulfates for all data collected during the entire study. These figures show the softener effluent met USPHS drinking water standards for these

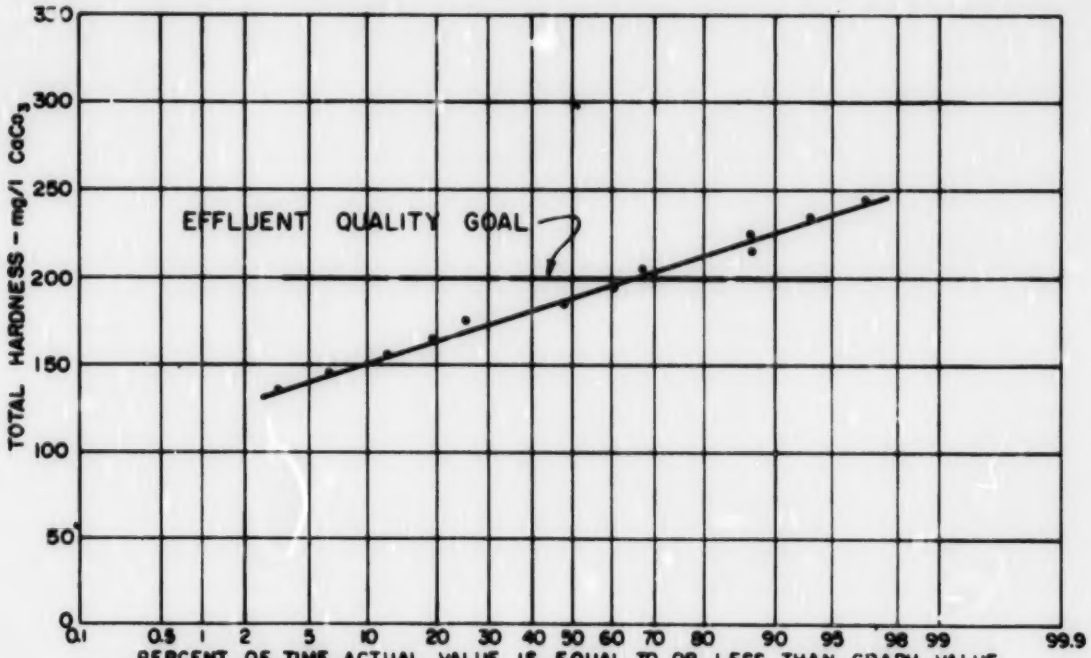

FIGURE 9 : SOFTENER EFFLUENT TOTAL HARDNESS INDUSTRIAL WATER USE QUALITY PROBABILITY PLOT

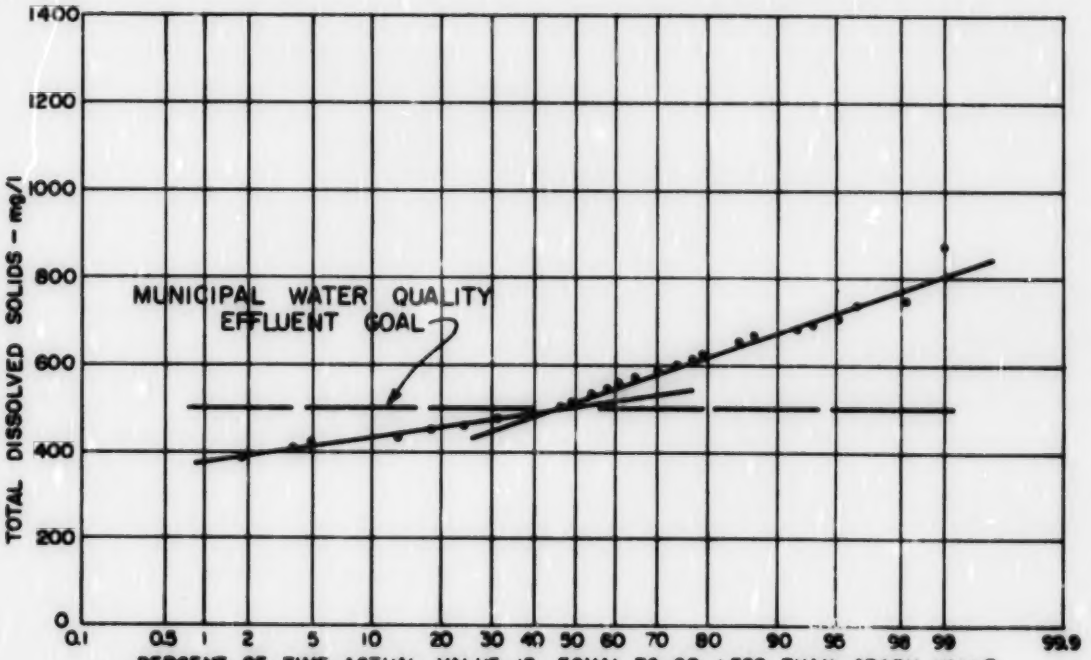
TOTAL DISSOLVED SOLIDS - mg/l
1400
1200
1000
800
600
400
200
0
MUNICIPAL WATER QUALITY EFFLUENT GOAL
0.1 0.5 1 2 5 10 20 30 40 50 60 70 80 90 95 98 99 99.9
PERCENT OF TIME ACTUAL VALUE IS EQUAL TO OR LESS THAN GRAPH VALUE

FIGURE 10 : SOFTENER EFFLUENT TOTAL DISSOLVED SOLIDS PROBABILITY PLOT

constituents most of the time irrespective of whether the operational target was municipal water supply quality or industrial water use quality No appreciable removal of sulfates were anticipated with the processes used. It should be pointed out again, however, that the raw water was not taken solely from Kittanning Run, the acid mine drainage impregnated stream, and hence, the raw water was lower in total dissolved solids and sulfates than had been anticipated.

SLUDGE PRODUCTION CONSIDERATION

Because the major emphasis in this study was directed towards effluent quality, chemical usage, operational considerations, and cost, only general data on sludge production was collected. The process flow diagram of the

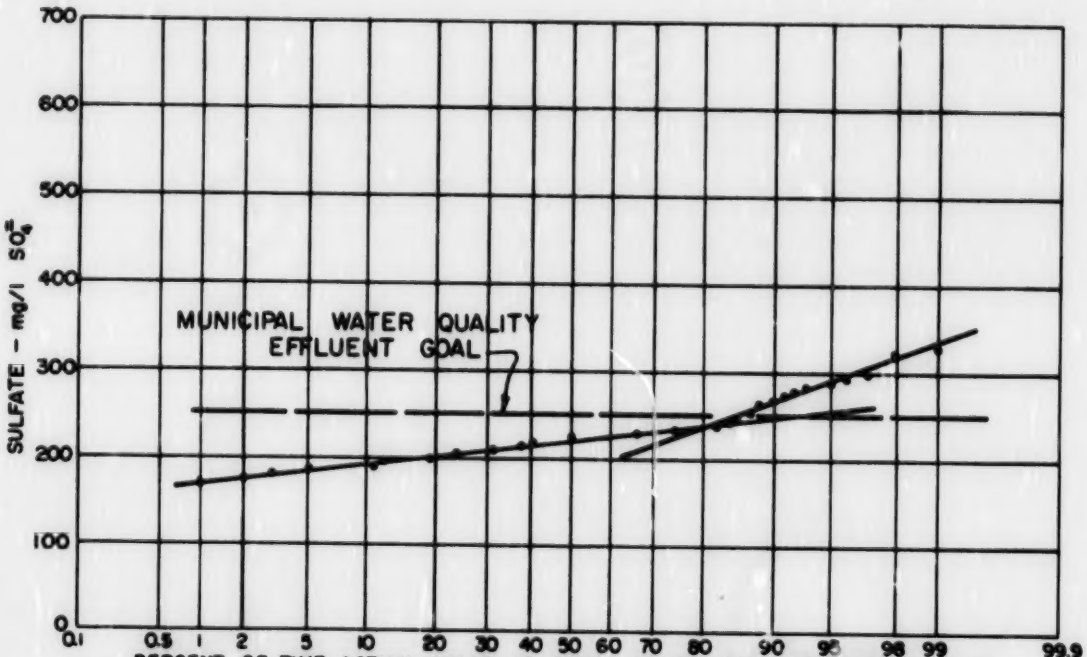

FIGURE II : SOFTENER EFFLUENT SULFATES PROBABILITY PLOT

plant shown in Figure 1 schematically illustrates the sludge handling procedures utilized. Underflow from both the neutralization stage sedimentation units and the Clariflow softener units was thickened in a gravity thickener. Thickened sludge was then vacuum filtered and the resulting sludge cake was landfilled.

It was decided that the best estimate of the amount of sludge produced and handled from the neutralization and softening stages could be obtained from a tally of the number of truck loads of filtered sludge hauled out for disposal. It was estimated from a limited amount of data on filter cake characteristics that each truck load contained approximately 2600 Kg (5700 lbs) of dewatered cake (approximately 25-percent total solids). A total of 390 truck loads of sludge were hauled out over the duration of the study

(Aug 12 - Dec. 8, 1974). The estimated sludge production from the overall process, therefore, was 690 kg (1520 lbs) of dewatered cake per 1000 cu m (.264 MG) of water processed. It was not possible to determine what fraction of the total sludge came from the neutralization stage and how much was generated in the softening operation. From the limited data available, vacuum filter yields ranged from approximately 37 kg (82 lbs) to 57 kg (126 lbs) of dewatered cake per square meter (10.8 ft) per hour. The sludge filtered well and no unusual operating problems were encountered.

CHEMICAL FEED CONSIDERATIONS

A complete discussion of the theoretical requirements for lime and soda ash for both the neutralization and softening stages is presented in Appendix A. All equation numbers used below refer to the chemical reactions presented therein. A summary of the chemical feed data is presented in Table 5.

The neutralization stage lime feed data presented in Table 5 was calculated in three different ways for comparison as follows:

1. Theoretical. The theoretical lime requirement is based on the chemical reactions (1), (2) and (4), assuming these proceed to completion.

2. Needed. The difference between the total metal ion concentration in the raw water and the dissolved metal ion concentration in the first stage sedimentation basin is multiplied by the appropriate lime demand calculated from the assumed reactions using equations (1), (2), (4), (6), and (7). The total "needed" lime demand is the sum of the demand for the individual reactions.

3. Observed. The observed lime demand calculations were based on the amount of calcium increase observed from data collected during the study.

Analysis and comparison of the results computed by the different methods, together with an observed decrease in sulfate concentration in the first stage, suggests that some calcium was removed from the first stage as a precipitated calcium sulfoaluminate in accordance with the following reaction:

$$CaO + CaSO_4 + 2\ Al(OH)_3 \xrightarrow{H_2O} 3\ CaSO_4 \cdot Al_2O_3 \cdot CaO \cdot XH_2O$$

This equation was used to correct the observed values calculated in accordance with method 3 above, based on observed sulfate loss to yield the corrected observed values in Table 5.

The softener lime feed was calculated similarly in three ways as follows:

1. Theoretical. The amount of lime required based on reactions (6) and (7).

2. Needed. The amount of lime required based on observed results

TABLE 5. SUMMARY CHEMICAL FEED DATA*

Period	Total Lime (mg/l CaO)			Neutralization Lime (mg/l CaO)				Softener Lime (mg/l CaO)			Softener Soda Ash (mg/l Na_2CO_3)				
	Mtr.	Thr.	Need.	Thr.	Need	Obs.	C.Obs.	Thr.	Need	Obs.	C-1	C-2	C-3	C-4	Mtr.
1974															
8/12-19	161	143	143	143	143	159	183	--	--	--	281	247	264	315	221
9/1-15	153	126	130	82	105	85	110	44	25	--	--	182	158	213	239
9/16-19	94	122	131	94	97	60	86	28	34	--	192	223	209	243	257
9/20/25	154	149	154	89	90	59	86	50	64	--	253	230	180	231	225
9/26-29	100	126	143	100	100	60	102	26	43	--	200	223	203	220	222
10/1-4	150	134	153	91	92	66	102	43	61	--	271	240	214	234	242
10/5-7	166	141	154	94	101	64	113	47	53	--	288	226	180	217	241
10/8-10	159	140	161	94	108	57	108	46	53	--	273	209	166	204	241
10/12-16	162	127	135	88	112	81	119	39	23	--	286	207	158	222	244
10/17-22	154	113	119	78	92	73	87	35	27	--	184	120	66	137	131
10/30-11/10	104	111	130	80	84	63	72	31	36	50	127	113	98	112	76
11/11-17	104	113	119	87	91	73	88	26	28	43	78	68	66	84	58
11/18-21	92	113	121	97	105	87	110	16	16	4	93	124	87	100	65
11/22-24	103	120	127	102	109	91	104	18	18	0	166	186	150	166	166
11/25-30	107	120	134	96	97	94	98	24	37	14	315	245	228	330	256
12/3-8	145	138	155	102	102	98	98	36	53	48	252	244	248	317	250
1975															
6/16-26				124	119	129							264	314	

*Definitions of Headings (all values reported in mg/l)
(1) Mtr.-- metered amount of lime or soda ash as recorded by plant operators.
(2) Thr.-- theoretical CaO requirement based on complete reactions.
(3) Need -- required amount of CaO based on observed results.
(4) Obs. -- amount of lime based on calcium increase.
(5) C.Obs. -- observed lime feed corrected for sulfate loss.
(6) C-1 -- soda ash feed calculated from metered lime feed
(7) C-2 -- soda ash feed calculated from theoretical lime feed required
(8) C-3 -- soda ash required based on observed results.
(9) C-4 -- soda ash feed based on sodium increase.

for free carbon dioxide, carbonate hardness, magnesium and manganese using equations (9), (12), and (7) respectively.

3. Observed. The observed lime demand calculations were based on the observed calcium increase in the mix box ahead of the softener.

Soda ash dosages were also calculated in several different ways as follows:

Method 1. The soda ash required based on the metered lime feed using equation (13).

Method 2. This method was the same as Method 1 except the theoretical lime requirement as defined above was used as the basis for calculation.

Method 3. The amount of soda ash required based on observed results using equations (14), (15), and (16).

Method 4. This method used the observed increase in sodium as the basis for calculation and hence is similar to the method defined as observed in the discussion on lime feed calculations.

Treatment plant records were used to determine the metered chemical usage for both lime and soda ash. For comparative purposes, the weighted mean values for the various methods of calculating the chemical feed requirements and the metered values are given in Table 6. These comparisons will be discussed in detail in the following section.

DISCUSSION

The following discussion will summarize observations made during the study and will concentrate on operational problems encountered. Table 7 presents summary performance data for the total study together with brief comments on operational observations.

Neutralization Stage

As has been indicated earlier, it was not possible to optimize the neutralization stage operation for the production of stream release quality water during the study. The collected data show the effluent requirements for stream release were easily met, however, it should be pointed out that the design capacity of the neutralization stage facilities is 28,400 cu m/day (7.5 MGD) although the actual flow during the study was only 13,200 cu m/day (3.5 MGD). It is not anticipated that there would be any difficulty in meeting the Department of Environmental Resource's requirements at the design flow even though the detention time in the first stage clarifiers is only 3.08 hours.

The light fragile hydrous iron floc formed in the slow mix and aeration basin was easily broken up as the water fell from the aeration basin effluent weir into the collection trough leading to the sedimentation basin. Floc

TABLE 6. COMPARISON OF WEIGHTED MEAN VALUES OF CHEMICAL FEED DATA

	Phase I				Phase II			
		Lime		Soda		Lime		Soda
	Total	Neutr.	Soft	ash	Total	Neutr.	Soft	ash
Metered (1)	143	--	--	239	111	--	--	87
Theoretical (2, 7)	137	98	39	220	113	85	28	113
Needed (3, 8)	147	106	41	219	121	92	29	111
Observed (4, 9)	--	87	31	253	--	72	36	114
Corrected observed (5)	--	113	--	--	--	85	--	--
Calc.-metered lime (6)	--	--	--	264	--	--	--	126

Note: All values are in mg/l.
(1) Meter -- metered amount of lime or soda ash as recorded by plant operators.
(2) Theor. -- theoretical CaO requirement based on complete reactions.
(3) Need -- required amount of CaO based on observed results.
(4) Observ. -- amount of lime based on calcium increase.
(5) Cor.Obs. -- observed lime feed corrected for sulfate loss.
(6) Calc. 1 -- soda ash feed calculated from metered lime feed.
(7) Calc. 2 -- soda ash feed calculated from theoretical lime feed required.
(8) Calc. 3 -- soda ash required based on observed results.
(9) Calc. 4 -- soda ash feed based on sodium increase.

TABLE 7. SUMMARY PERFORMANCE DATA

Period 1974	pH			Hardness		Dissolved Manganese			T-oC.	Comments
	Sed.	Clar.	Recarb.	Clar.	Recarb.	Sed.	Clar.	Well	Clar.	
8/12-19	11.0	10.8	9.2	119	87	.01	.01	--	21	No softener reaction zone solids, poor clarification, but objective met.
9/1-15	9.5	11.0	10.8	141	133	.06	.01	--	18	Poor clarification - final hardness in excess of objective (100 mg/l). Alum = no help.
9/16-19	7.9	10.7	10.2	119	109	2.0	.26	.16	18	Poor clarification, manganese above recommended limit.
9/20-25	7.5	11.2	9.1	128	112	3.6	.16	.16	15.6	Good floc with good settling at pH 11. Mn too high. Sed. Basin = heavy floc carryover
9/26-29	7.2	10.6	7.4	112	110	4.2	.23	.30	15.7	Poor clarification. Mn too high.
10/1-4	7.4	11.4	--	112	--	4.0	.08	--	12.1	Excellent softener operation. Aeration controlled to reduce floc carryover from sed.basin
10/5-7	7.3	11.4	--	127	--	3.3	.04	--	12.5	Excellent softener operation. Aeration controlled and coag. aid added to sed. basin (1mg/l).*
10/8-10	7.3	11.3	--	123	--	3.9	.27	--	12.5	Excellent softener operation Mn too high. Potassium Permanganate feed = 6.7 mg/l Coa. Aid = 7.5 mg/l
10/12-16	9.8	11.2	--	126	--	.08	.01	--	13.6	Poor clarification = non-settleable floc

TABLE 7 (continued)

Period 1974	pH			Hardness		Dissolved Manganese			T-°C.	Comments
	Sed.	Clar.	Recarb.	Clar.	Recarb.	Sed.	Clar.	Well	Clar.	
10/17-22	9.3	11.5	--	207	--	.10	.02	--	10	Poor softener operation. Tried coag. aid (1 mg/l) CA 253 = no improvement
10/28-11/10	8.6	11.2	7.8	182	176	.08	.08	.03	11	No reaction zone solids, poor clarification.
11/11-18	8.5	11.1	8.2	221	216	.16	.03	.02	6.4	Poor softener clarification. Feeding 10 mg/l Potassium permanganate I.D. line. Filter runs = 54 hr
11/19-21	8.5	10.5	8.0	218	212	.04	.03	.04	5.3	Softener clarification poor Brown floc carry-over. Filter runs = 30 hr
11/22-24	8.5	10.8	8.3	159	165	.36	.04	.04	5.0	Softener clarification poor; product water quality ok. Filter runs = 20 hr
11/25-30	8.7	11.2	8.7	112	103	1.2	.08	.08	2.3	Poor softener clarification Filter runs = 20 hr
12/3-8	7.6	11.4	8.3	133	168	4.8	.37	.19	1.5	Poor softener clarification Filter runs = 20 hr

*Floc carry-over to the softener stage was reduced, but this had no apparent effect on softener operation.

breakup undoubtedly resulted in reduced settling basin efficiency which is very important if stream release quality water is being produced. As indicated earlier, some modification in operating conditions did not appreciably reduce the amount of iron floc carryover from the sedimentation basins.

Addition of a cationic polymer (Calgon 233) to the sedimentation basin influent, however, significantly improved settling and resulted in a more dense sludge. Optimization of the first stage sedimentation step by evaluating different coagulant aids could be done in a separate study prior to commencing operation for the sole purpose of meeting stream release quality objectives.

The primary function of the neutralization stage during this study was to remove iron and manganese prior to the softening step. Iron removal did not present any particular problems as evidenced by the stream release quality data analysis. In fact, some carryover of iron floc from the first stage to the softener may have improved clarification in the softening step. On the other hand, manganese oxidation is highly dependent upon pH and the USPHS limits on manganese were not achieved during portions of the study. Four pre-softening stage conditions were investigated to determine the manganese removal capabilities of the system. Summary results for the different control conditions were as follows:

Control condition	Manganese, mg/l	
	Total	Dissolved
Neutralization to pH 7.0	3.90	2.80
Neutralization to pH 9.0	.50	.08
Neutralization to pH 11.1	.20	.01
Neut. to pH 7.0 (10 mg/l $KMnO_4$)	.86	.10

The addition of potassium permanganate to the raw water influent line ahead of the neutralization stage is recommended in order to consistently assure oxidation of the manganese and its subsequent removal.

Magnesium removal in the first stage also was highly dependent upon pH as expected. Very little reduction in raw water magnesium (average value approximately 18 mg/l) was noted when the first stage pH was set at 7.0. Operation of the first stage at pH 9.0 to 9.5 resulted in a magnesium reduction of approximately 50 percent whereas operation at a pH 11.0 to 11.5 resulted in removal of nearly all of the magnesium.

Lime addition to the neutralization process did not contribute an equivalent amount of calcium hardness in the sedimentation basin effluent. Some calcium (approximately 15 mg/l) was apparently precipitated as calcium sulfate, as further evidenced by a comparable sulfate loss in the neutralization stage. This effect is of interest since it would tend to reduce effluent sulfate values and this would be particularly desirable for potable water use.

Softening Stage

Successful operation of a solids-contact softening unit of the type employed in this study requires the recycle of settled solids back to the reaction zone for intimate mixing of previously formed floc with the applied flow and added chemicals. The manufacturer's instructions called for controlling the slurry concentration at 5 to 10 percent by weight in the bottom of the basin in order to achieve optimal process operation (11). Except for a period of about three weeks during the study, however, it was not possible to maintain the slurry at the desired concentration. The general appearance of the softener effluent, for the most part, could be characterized as a turbid water containing a finely divided precipitate. Generally, samples drawn from various levels and zones within the softening unit failed to show any differential in the solids concentration which would indicate the desired solids buildup was occurring. It was possible to build the reaction zone solids up to the desired levels for brief periods when the pH in the softener was approximately 11.3 and the temperature of the water being treated was 12°C or higher. It was not possible to obtain good clarification when water temperatures were below 12°C. Likewise, good clarification was not obtained at pH 9.6 even when the water temperature was approximately 17°C. The reason for the good clarification only at the extremely high pH (11.3±) is not understood but it may be due to the virtually complete precipitation of magnesium as the hydroxide at that pH.

Even when good clarification was being achieved, the municipal water supply hardness objective of 100 mg $CaCO_3$/l was not realized with the desired consistency. Several possible reasons for this higher-than-expected effluent hardness exist. Temperature-time relationships are significant in lime-soda ash softening. The theoretical reaction zone retention time for the softening unit used in the study was 25 minutes although actual retention times may have been less than that. No definitive retention time studies were made during the project, however, which would permit specific conclusions on this point. Assuming the actual retention time in the reaction zone was approximately 25 minutes, water temperatures should be at least 10°C in order to achieve the desired effluent hardness of 100. The following table shows the influence of temperature and time on effluent hardness for given chemical dosages:

TABLE 8. RESIDUAL HARDNESS OBSERVED AT DIFFERENT TEMPERATURES
(mg/l as $CaCO_3$)

Time (min.)	Temperature - °C		
	2	10	20
30	124	84	64
60	82	71	46
90	71	67	46
120	75	62	--

Water temperatures of less than 10°C would be experienced in the plant studied commonly during winter months when Kittanning Run is being used as a source and hence, slow reaction times in the softener would make it impossible to achieve the desired effluent hardness of 100 mg $CaCO_3$/l at the design flow of 13,200 cu m/day (3.5 MGD) per unit. Impoundment of this water source would tend to moderate and stabilize the temperature somewhat and should result in improved performance of all the treatment units.

Sulfate ion has been shown to have a major effect in slowing the rate of reaction and formation of calcium carbonate when less than 1000 mg/l (0.1%) of preformed calcium carbonate is present (12). The water being treated in this study contained relatively high sulfates (See Fig. 12) and no bicarbonates. Hence, all the initial hardness to be removed was non-carbonate hardness (NCH). As indicated above, it was virtually impossible to maintain the desired level of reaction zone solids in the softener unit, even with internal sludge recycle, so that the condition necessary to effect optimum softening with this high sulfate water simply did not exist.

Furthermore, according to Symons (13), formation of complex magnesium salts may be a problem with waters containing high NCH. These complex salts are more soluble than magnesium hydroxide; hence, theoretical hardness reduction may not be obtainable. The soda ash typically used to remove NCH in the lime-soda ash process tends to cause formation of basic carbonate complex salts such as $Mg_4(OH)_2(CO_3)_2$.

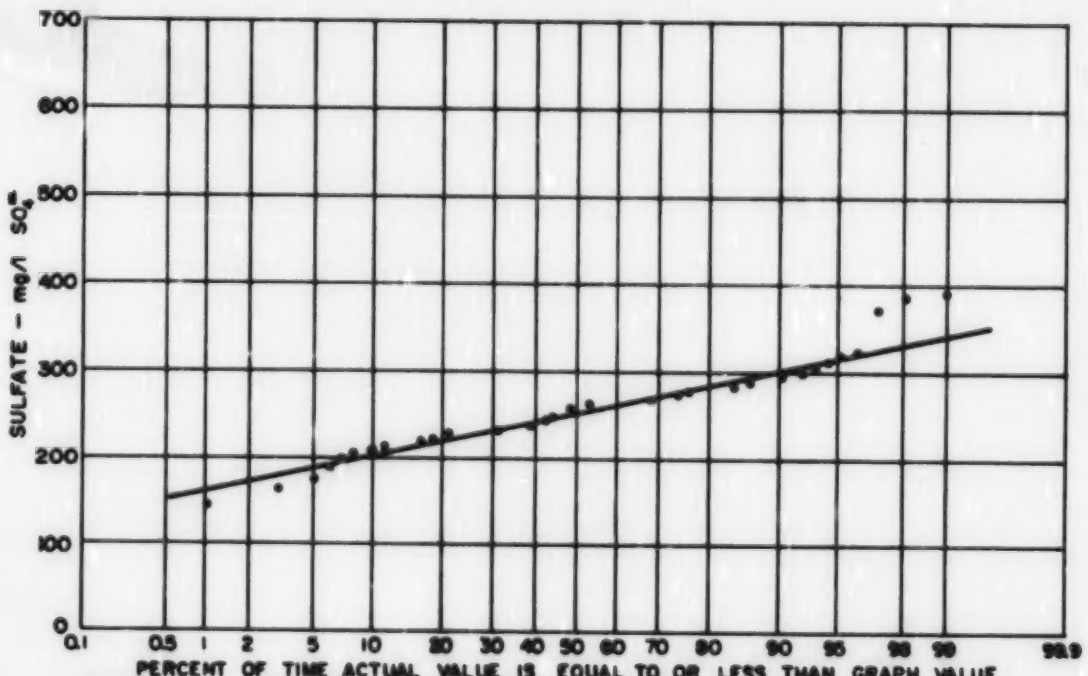

FIGURE 12: RAW WATER SULFATES PROBABILITY PLOT

Under some conditions (temperature, presence of anions, complex salt formation, etc.) the precipitated salts formed in the softener may remain in collodial form. This observation may account partially for the observed difficulty in obtaining good clarification during most of the study. In this study however, use of alum and polymers did not noticeably improve softener operation.

Chemical Utilization

Using the weighted mean chemical feed data presented in Table 6, it is possible to make some observations concerning chemical utilization efficiency (CUE) for the operating conditions of the study. For the purposes of these observations, chemical utilization efficiency is defined as:

$$CUE(\%) = \frac{\text{Observed chemical usage}^{a}}{\text{Theoretical chemical requirement}^{b}} \times 100$$

a) See definitions 5 and 9, Table 5
b) See definitions 3 and 7, Table 5

Table 9 summarizes the results of the chemical utilization efficiency for the two phases of the study.

TABLE 9. CHEMICAL UTILIZATION EFFICIENCY
(Percent)

Process Stage	Phase I		Phase II	
	Lime	Soda Ash	Lime	Soda Ash
Neutralization	77.0	--	100.	---
Softening	79.5	115.	128.	99.1

The results from Phase I indicate that the neutralization of the influent water together with the desired pH adjustment for softening only required approximately 80 percent of the theoretical lime requirement. On the other hand, the soda ash requirement was approximately 15 percent greater than the theoretical demand. This observation is particularly significant as it shows that even with a 15-percent excess of soda ash, it was not possible to achieve the desired effluent hardness of 100 mg $CaCO_3$/l. Possibly, if this excess were increased, some additional hardness removal could be achieved but it could increase the chemical costs significantly.

The Phase II data indicates that the observed chemical usage was essentially equal to the theoretical requirements except of the lime utilization in the softening stage. No explanation of the reason for the considerably higher lime use in the softening stage was apparent in the data.

Figure 13 shows the relationship between softener effluent hardness and the soda ash dosage for the data collected during the study. Although there is some scatter in the plotted points, the trend lines shown indicate the minimum attainable hardness under the conditions of the study was approximately 120 mg $CaCO_3$/l regardless of the soda ash dosage. As indicated earlier, waters containing appreciable amounts of NCH are difficult to soften and the data from this study support this observation.

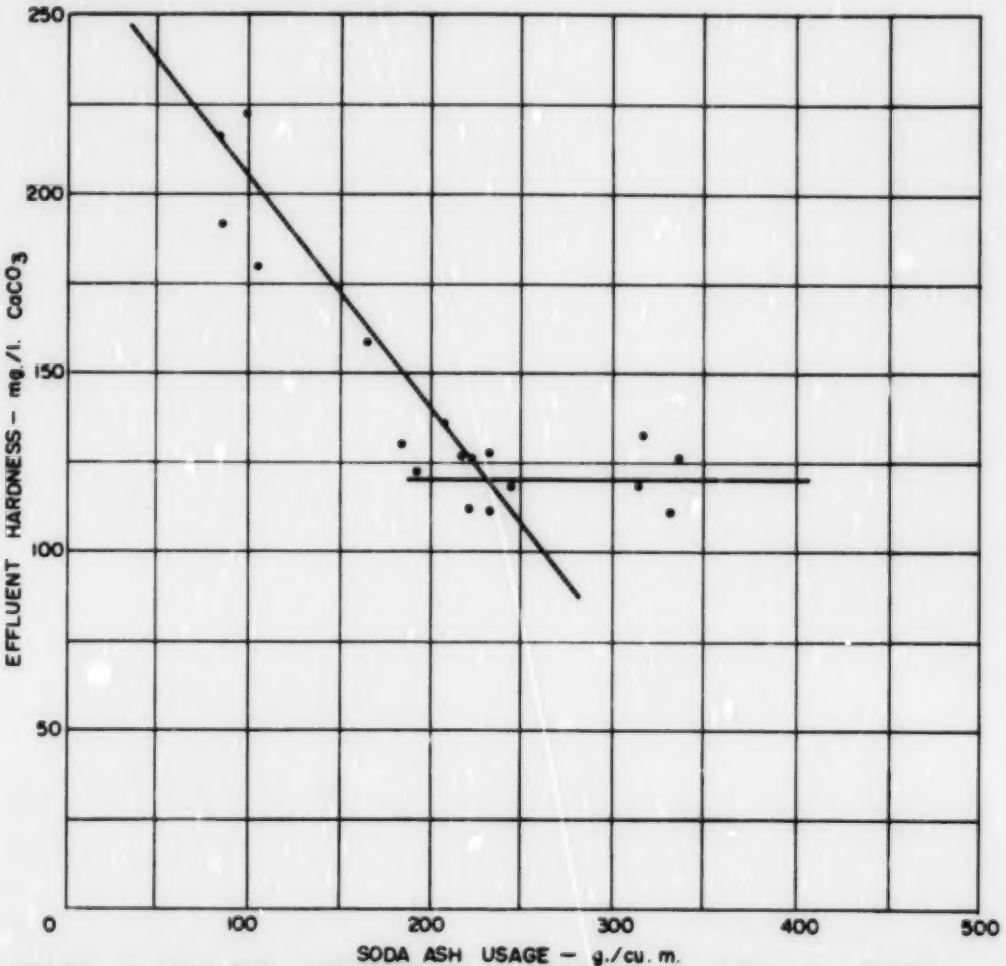

FIGURE 13 : EFFLUENT HARDNESS AS A FUNCTION OF SODA AS USAGE

Figure 14 shows the raw water lime demand required to achieve a given pH. the additional lime dosage required to raise the pH to 11.3, the optimum point for best clarification in this study, amount to 21 g/cu m (173 lbs/MG) when compared to that required to raise the raw water pH to 10.6, a common target pH for magnesium removal. If compared to a common target pH of 9.6 for calcium removal, this required differential is 37 g/cu m (311 lbs/MG). It was impossible to determine whether the high pH required during the study was due to the raw water characteristics only or was due to a combination of the water to be treated and the softening unit used. It is felt, however, that it is primarily related to the high NCH water source, typical of acid mine waters.

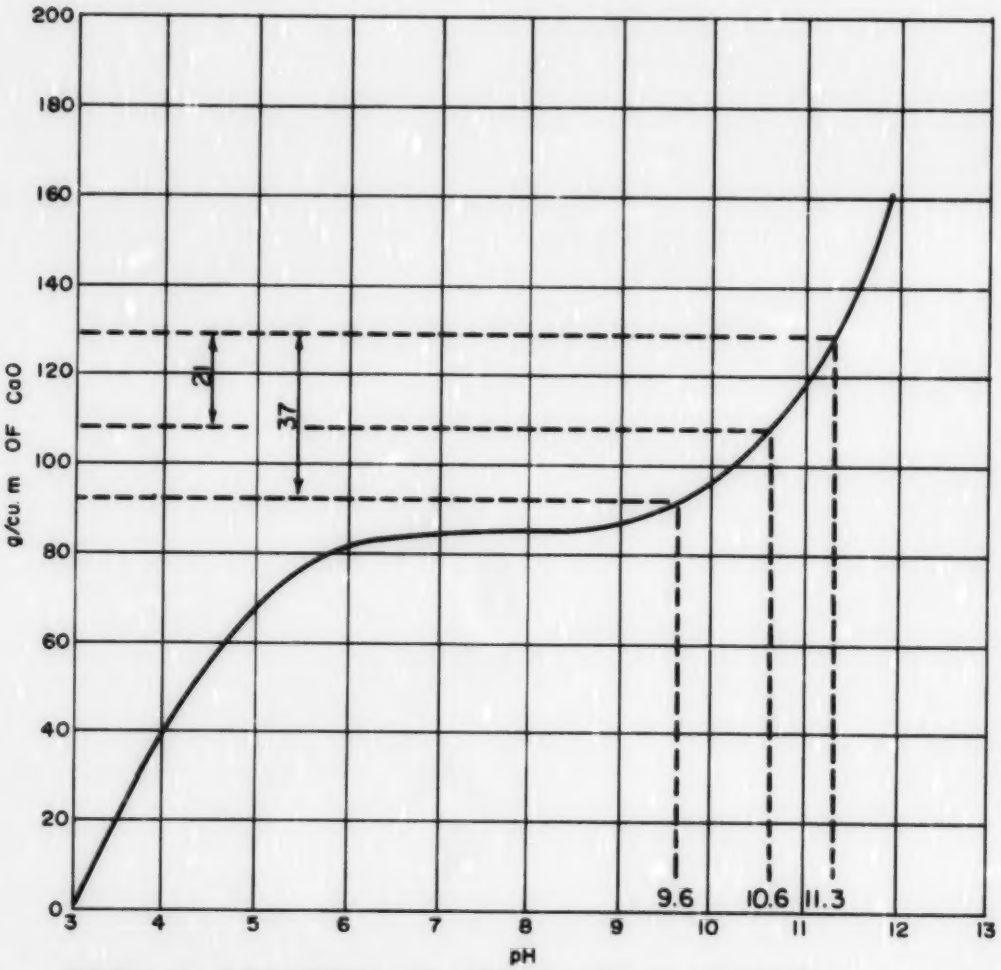

FIGURE 14 : RAW WATER LIME DEMAND FOR pH INCREASE

The data and curves previously presented reflect operating conditions and results for the particular blend of waters from Impounding Dam and Kittanning Run used in the study. Different raw water sources or an appreciable change in the ratio of I.D. to K.R. would likely change the results obtained. Any such waters would probably be difficult to soften to low hardness.

Filter Operation

It is unfortunate that it was only possible to obtain filter operating data for approximately one month during the study. Data collected during this period generally indicated satisfactory performance with filter runs of 20 to 54 hours being observed (See Table 7). Filtration had little observable effect on hardness reduction but in several instances, it did significantly improve manganese removal. It was not possible, with the limited duration of the filtration study, to determine the long term effects of the poor clarifier performance on possible calcium carbonate buildup on the filter media, etc.; nor was it possible to fully evaluate the effect of filter "seasoning" on the removal of iron and manganese.

One major operating problem developed in the period October 16 to 22 when the filters first were placed in operation during a time when a polymer (Calgon 253) was being used in an attempt to improve softener clarification. The filters "fouled" very rapidly with a gelatinous material that coated the media. The filters were taken out of service and a variety of measures were taken to restore the media when normal backwash procedures failed to clean the media. The filters were again placed in operation after cleaning and operated for approximately one month with no evidence of fouling. It is believed the filter clogging, therefore, was directly a result of the polymer addition.

SECTION VI

ECONOMIC EVALUATION

CAPITAL INVESTMENT COSTS

The complete water supply system consists of at least three main cost elements which must be considered in an economic evaluation. They are: 1) the production of raw water, 2) treatment, and 3) transmission of treated water to storage and distribution. Only the second of these items, treatment of raw water, will be addressed in this report from the standpoint of capital, operating and maintenance costs. The type of treatment relative to producing a finished water of stream release quality is simply lime neutralization, followed by mixing, aeration and settling. Costs for producing a higher grade of effluent will be based on a treatment process consisting of iron removal (neutralization), manganese removal, softening, and, in the case of a municipal supply, chlorination.

Capital investment costs required to provide the unit operations (concrete tankage, mechanical equipment, instrumentation, etc.) and control building necessary for the treatment described above, cannot be directly related to the total cost of the Altoona facility. The principal reason for this is that the nearly $5 million expenditure for the Altoona plant includes the capacity to treat 15 mgd through the neutralization stage, in addition to 7.0 mgd in the softening stage. In order to resolve this inconsistency and include the initial dollar investment for a treatment plant pertinent to our economic evaluation, the actual costs of the Altoona plant were adjusted to reflect the capital cost for a 7.0 mgd facility (neutralization through softening). Conversion to 1975 dollars was accomplished by using a suitable factor based on construction cost indexes for 1970 (actual plant cost) and 1975 as reported by the Engineering News-Record. Accordingly, the capital outlay required to provide the treatment levels discussed in this report would be $1.8 million for the neutralization plant, and $3.8 million for the neutralization/softening plant. Land and rights-of-way costs are not included and, as previously mentioned, costs involving water source development, transmission mains, storage and distribution systems are not considered. By way of explanation, with regard to the rather high cost of the neutralization plant, it should be pointed out that a more sophisticated system than might be required was provided because of the two-stage nature of the treatment process and the need for more refined operational control.

Annual payments for these two facilities over twenty-five (25) years at 7% interest amount to $154,440 for the neutralization plant, and $326,040.00 for the neutralization/softening plant - or approximately 1.6¢/ cu m (6¢/1000 gallons) and 3.4¢/cu m (12.7¢/1000 gallons) respectively.

OPERATING AND MAINTENANCE COSTS

Personnel, chemicals and power were the major items of cost considered under operating expense. Personnel costs were based on the actual manpower requirements at the Altoona plant. The workforce needed to operate a two-stage process (neutralization and softening) consisted of a plant supervisor, four plant operators, four assistant operators and 1/2 chemist. The total annual cost for personnel was determined to be $135,200.00, or based on water production, about 1.4¢/cu m (5.2¢/1000 gallons). The personnel costs for operating a plant strictly for neutralization to meet a stream release discharge quality would not be nearly as much. In our judgment this operation could be adequately handled by two, or possibly three people at a cost of not more than $50,000.00/year. In this case the personnel operating expense is about .5¢/cu m (2¢/1000 gallons).

Chemical and electrical costs were evaluated in terms of the chemicals and power required to produce a water quality meeting stream release standards, and a 120 mg/l or 200 mg/l hardness objective. As one might expect, the chief difference in chemical cost associated with these objectives is in the quantity of soda ash used. A summary of the power, chemical and maintenance costs is presented in Table 10.

SUMMARY OF TREATMENT COSTS

Capital investment costs and operator personnel expenses will be fixed costs under each of the two plants considered, i.e., either a 7.0 mgd neutralization plant alone, or a 7.0 mgd neutralization/softening plant. Chemicals power and maintenance costs will vary depending on the finished water quality objective (assuming raw water quality to be the same). Table 11 summarizes these costs.

TABLE 10. SUMMARY OF COST DATA--POWER, CHEMICALS, AND MAINTENANCE

Chemicals	Neutralization only stream release quality	Neutralization and Softening Hard(120mg/l)	Hard(200mg/l)
Lime			
g/cu m	99	165	120
lb/1000 gal	.83	1.38	1.00
¢/cu m	0.32	0.51	0.37
¢/1000 gal	1.20	1.93	1.40
Soda Ash			
g/cu m	--	224	108
lb/1000 gal	--	1.87	0.90
¢/cu m	--	1.63	0.78
¢/1000 gal	--	6.17	2.97
Pot. Perm.			
g/cu m	--	10	10
lb/1000 gal	--	0.08	0.08
¢/cu m	--	1.14	1.14
¢/1000 gal	--	4.32	4.32
CO_2			
g/cu m	--	51	51
lb/1000 gal	--	0.43	0.43
¢/cu m	--	0.49	0.49
¢/1000 gal	--	1.85	1.85
Chlorine			
g/cu m	--	1	1
lb/1000 gal	--	.008	.008
¢/cu m	--	0.37	0.37
¢/1000 gal	--	1.40	1.40
Total chemical cost			
¢/cu m	0.32	4.14	3.15
¢/1000 gal	1.20	15.67	11.94
Power usage, KWH/day	1600	2000	2000
Power cost @ $.027/KWH			
¢/cu m	0.32	0.41	0.41
¢/1000 gal	1.23	1.54	1.54
Maintenance cost			
¢/cu m	0.26	0.52	0.52
¢/1000 gal	0.98	1.96	1.96
Total cost, Chemicals Power and Maintenance			
¢/cu m	0.90	5.07	4.08
¢/1000 gal.	3.41	19.17	15.44

TABLE 11. SUMMARY OF TREATMENT COSTS

	Neutralization Only	Neutralization and Softening	
	stream release quality	Hard(120mg/l)	Hard(200mg/l)
Capital Investment			
¢/cu m	1.58	3.35	3.35
¢/1000 gal	6.00	12.70	12.70
Operating and Maintenance			
¢/cu m	1.43	6.43	5.45
¢/1000 gal	5.41	24.37	20.64
Total Cost			
¢/cu m	3.01	9.78	8.80
¢/1000 gal	11.41	37.07	33.34

SECTION VII

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SECTION VIII

APPENDICES

APPENDIX A. THEORETICAL CONSIDERATIONS : CHEMICAL FEED DATA

Calculation of Neutralization Lime Need

Two general processes will cause a lime demand during the neutralization stage. The first process is the neutralization of the free acid present.

(1) $H_2SO_4 + CaO \longrightarrow CaSO_4 + H_2O$

This reaction shows that one mole of lime will neutralize two moles of free acid. Therefore at a pH of 3.0, the average pH of the raw water, the amount of lime required to neutralize the free acid would be 28.0 mg/l CaO.

The second process that will exert a lime demand is the neutralization of the acid produced by the hydrolysis of the polyvalent metal ions present in the raw water. The reactions and corresponding lime demand can be summarized as follows:

(2) $Fe_2(SO_4)_3 + 3\ CaO \xrightarrow{H_2O} 2\ Fe\ (OH)_3 + 3CaSO_4$ 1 mg/l Fe^{+3} requires 1.51 mg/l CaO

(3) $FeSO_4 + CaO \xrightarrow{H_2O} Fe(OH)_2 + CaSO_4$ 1 mg/l Fe^{+2} requires 1.00 mg/l CaO

(4) $Al_2(SO_4)_3 + 3CaO \xrightarrow{H_2O} 2\ Al(OH)_3 + 3CaSO_4$ 1 mg/l Al^{+3} requires 3.12 mg/l CaO

(5) $2\ Al(OH)_3 + CaO + SO_4^{-2} \xrightarrow{H_2O} 2\ AlO_2^- + CaSO_4$ 1 mg/l Al^{+2} requires 1.04 mg/l CaO

(6) $MgSO_4 + CaO \xrightarrow{H_2O} Mg(OH)_2 + CaSO_4$ 1 mg/l Mg^{+2} requires 2.31 mg/l CaO

(7) $MnSO_4 + CaO \xrightarrow{H_2O} Mn(OH)_2 + CaSO_4$ 1 mg/l Mn^{+2} requires 1.02 mg/l CaO

Theoretically in order for precipitation to occur the ion product, [Metal][OH^-], must exceed the solubility product, Ksp, for the compound. From solubility products, a graph of polyvalent metal ion concentration at various pH's has been constructed (see Figure A-1).

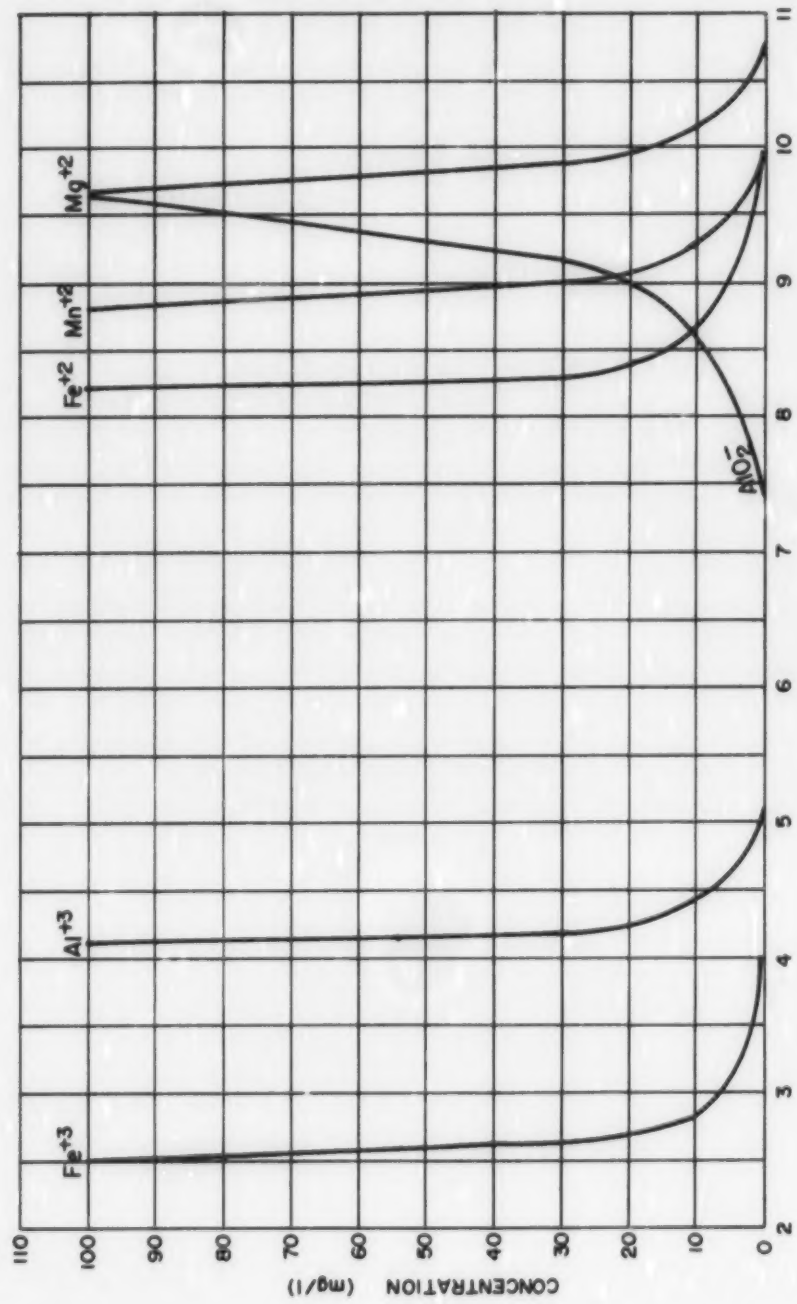

FIGURE A-1 : POLYVALENT METAL ION CONCENTRATION AT VARIOUS pH'S

This graph shows that precipitation of ferric iron will occur completely when the pH is raised to 4 or above during the neutralization stage. The precipitation of ferrous hydroxide does not occur until a much higher pH is attained. However, in the raw water ferric iron predominates, therefore equation (3) and its corresponding lime demand is insignificant.

Aluminum hydroxide precipitation (4) is essentially complete when the pH is above 5, however, aluminum is amphoteric. When the hydroxide ion [OH^-] concentration is elevated, the aluminum reacts further (5) and will cause an additional lime demand. This demand is slight because of insignificant aluminum concentrations in the sed. basin and can be ignored in all test periods except for August 12-19.

Theoretically, the precipitation of manganous hydroxide and magnesium hydroxide in the raw water will occur at pH's of 9.5 and 10.0 respectively and will be essentially complete at pH's of 10.1 and 10.8.

The neutralization lime feed presented in Table 5, Summary Chemical Feed Data (see Page 26) was calculated three ways.

1. Theoretical - the theoretical lime requirement is based only on reactions (1), (2), and (4) assuming these proceed to completion.

2. Needed - the required amount of lime based on observed results. The difference between the total metal ion concentration in the raw water and the dissolved metal ion concentration from the sedimentation basin is multiplied by the appropriate lime demand Equations (1), (2), (4), (6), and (7) as needed. These demands are then added to get the total needed lime demand.

3. Observed - amount of lime needed based on calcium increase.

The observed lime feed is lower in most cases than the needed lime feed. We believe this is due to lime-calcium sulfate reactions with precipitated aluminum hydroxides as further evidenced by sulfate loss in the neutralization stage.

(8) $CaO + 3CaSO_4 + 2Al(OH)_3 \xrightarrow{H_2O} 3CaSO_4 \cdot Al_2O_3 \cdot CaO \cdot XH_2O$

When this reaction is taken into account, the corrected observed and needed lime feeds compare more favorably.

Calculation of Softener Lime Need

Lime required on the first stage of softening is related to free carbon dioxide, carbonate hardness, manganese, Mn^{+2}, and magnesium, Mg^{+2}, contents. The free carbon dioxide reaction and lime demand is

(9) $CO_2 + CaO \longrightarrow CaCO_3$ 1 mg/l CO_2 requires 1.27 mg/l CaO

For carbonate hardness the reactions are

(10) $Ca^{+2} + 2HCO_3^- + CaO \xrightarrow{H_2O} 2CaCO_3$

(11) $Mg^{+2} + 2HCO_3^- + CaO \xrightarrow{H_2O} CaCO_3 + Mg^{+2} + CO_3 =$

Since carbonate hardness and alkalinity are numerically equal when expressed as $CaCO_3$ the calculation for its lime requirement would be

(12) Alkalinity (Sed. basin in mg/l) x 0.56 = mg/l CaO

Manganese and magnesium reactions and lime demands would be the same as in the proceeding section -- equations (6) & (7).

The softener lime feed was calculated three ways.

1. Theoretical - amount of lime needed based on reactions (6) & (7)

2. Needed - the required amount of lime based on observed results. The lime demand due to the free carbon dioxide (9), carbonate hardness (12), magnesium (6) and manganese (7) contents.

3. Observed - the lime feed calculated from calcium increase in the mix box.

Calculation of Softener Soda Ash Need

Soda ash is added to the mix box effluent to remove noncarbonate calcium hardness. Precipitation of calcium as calcium carbonate occurs according to the reaction:

(13) $Ca^{+2} + Na_2CO_3 \longrightarrow CaCO_3 + 2Na^+$

For precipitation to occur, the pH must be about 9.5 or above. The soda ash demand is 2.64 mg/l Na_2CO_3 for each mg/l Ca^{+2} precipitated.

If insufficient lime is added in the first stage of softening, the manganese and magnesium ions in solution will react with the soda ash before the soda ash can react with the calcium. The equations are:

(14) $Mn^{+2} + 2Na_2CO_3 \xrightarrow{H_2O} Mn(OH)_2 + 2HCO_3^- + 4\ Na^+$

(15) $Mg^{+2} + 2Na_2CO_3 \xrightarrow{H_2O} Mg(OH)_2 + 2HCO_3^- + 4\ Na^+$

The soda ash demands for the above reactions are:

1 mg/l Mn^{+2} precipitated requires 3.86 mg/l Na_2CO_3

1 mg/l Mg^{+2} precipitated requires 8.72 mg/l Na_2CO_3

The softener soda ash feed was calculated two ways:

1. Needed - amount of soda ash required for observed results. In most cases this was calculated by the following formula:

(16) $(S.B. - Soft + SLF) \times 2.64 = N$

Where

S.B. = sed. basin effluent - total Ca value
Soft = softener effluent - dissolved Ca value
SLF = softener lime feed in mg/l Ca
N = needed soda ash

For 5 test periods; 9/16-19, 9/26-29, 11/18-21, 11/22-24, and 11/25-30 insufficient lime was added in the first stage of softening. Therefore, equations (14) and (15) with their soda ash demand were used in addition to the soda ash needed to precipitate the noncarbonate calcium hardness.

2. Observed - the soda ash feed calculated from the sodium increase in the reaction clarifier.

APPENDIX B. SUMMARY OF OPERATIONAL DATA

PERIOD: August 12 - 19, 1974

PROCESS OBJECTIVES: Neutralization Stage pH = 11.0
Softening Stage pH = 11.0
Hardness Objective = 100 mg/l

CHEMICAL FEED: Lime, Neutralization stage, kg/cu m = 0.183
Lime, softening stage, kg/cu m = 0
Soda Ash, soft stage, kg/cu m = 0.315
Carbon Dioxide, recarb., kg/cu m = 0.051

REMARKS: No build up of solids in softener reaction zone. Clarification zone turbid with finely divided precipitate present. Average operating temperature 21°C. No filter operation.

AVERAGE PERFORMANCE DATA*

Parameter	Raw water	Sed. basin effluent	Softener effluent	Recarb. effluent
Flow, mgd.	3.5	3.5	3.5	3.5
Flow, cu m/day	13,248	13,248	13,248	13,248
pH Units	3.0	11.0	10.8	9.2
Acid, tot.	173	0	0	0
Alk, tot.	0	105	109	74
Alk, "P"	0	85	84	19
Ca, tot.	27.7	141	41	33
Ca, diss.	--	132	42	32
Mg, tot.	17.7	1.1	0.4	0.4
Mg, diss.	--	0.4	0.3	0.3
Fe, tot.	17	0.8	0.3	0.3
Fe, diss.	--	0.2	0.1	0.1
Mn, tot.	4.5	0.2	.05	.06
Mn, diss.	--	0.1	.01	.01
Al, tot.	13.4	2.7	2.5	2.4
Al, diss.	--	2.5	2.5	1.4
Na	1.8	1.8	138	136
Hard. (calc.)	255	358	119	87
SO_4	273	242	223	220
Sp. C. umhos/cm^2	821	929	1010	785
TSS	--	--	--	--
TDS	--	--	--	--

*Unless otherwise noted, all results are expressed in mg/l of the constituent indicated. Dissolved values were determined on samples filtered through a borosilicate microfiber glass filter. Acidity, alkalinity, and hardness are expressed as calcium carbonate equivalent values.

PERIOD: September 1 - 15, 1974

PROCESS OBJECTIVES: Neutralization Stage pH = 9.5
Softening Stage pH = 10.5 - 11.0
Hardness Objective = 100 mg/l

CHEMICAL FEED: Lime, neutralization stage, kg/cu m = 0.11
Lime, softening stage, kg/cu m = 0.044
Soda Ash, soft stage, kg/cu m = 0.213
Alum, soft stage, kg/cu m = 0.02
Carbon Dioxide, recarb., kg/cu m = 0.096 (9/2 & 9/4 only)

REMARKS: No softener reaction zone solids build up. Clarity of softener effluent poor. Alum feed (20 mg/l) did not help. Recarb. pH could not be reduced to much under 10.0 with maximum CO_2 feed. Average operating temperature 18°C. No filter operation.

AVERAGE PERFORMANCE DATA*

Parameter	Raw water	Sed. basin effluent	Softener effluent	Recarb. effluent
Flow, mgd.	3.5	3.5	3.5	3.5
Flow, cu m/day	13,248	13,248	13,248	13,248
pH units	3.0	8.6	10.8	10.0
Acid, tot.	159	--	--	--
Alk, tot.	--	17	80	73
Alk, "P"	--	2	62	39
Ca, tot.	31	92	79	76
Ca, diss.	--	91	50	48
Mg, tot.	18	8.6	8.0	1.6
Mg, diss.	--	7.5	0.9	0.9
Fe, tot.	12.4	1.5	.28	.17
Fe, diss.	--	0.1	0.1	.1
Mn, tot.	4.7	.45	.08	.06
Mn, diss.	--	.06	.01	.01
Al, tot.	11.4	3.0	3.1	2.4
Al, diss.	--	1.9	2.2	1.7
Na	2.1	2.4	95	94
Hard. (calc.)	245	269	141	133
SO_4	242	211	260	211
Sp. C. umhos/cm^2	823	592	885	679
TSS	--	--	--	--
TDS	--	--	--	--

*Unless otherwise noted, all results are expressed in mg/l of the constituent indicated. Dissolved values were determined on samples filtered through a borosilicate microfiber glass filter. Acidity, alkalinity, and hardness are expressed as calcium carbonate equivalent values.

PERIOD: September 16 - 19, 1974

PROCESS OBJECTIVES: Neutralization Stage pH = 7.0
Softening Stage pH = 10.5 - 11.0
Hardness Objective = 100 mg/l

CHEMICAL FEED: Lime, neutralization stage, kg/cu m = 0.086
Lime, softening stage, kg/cu m = 0.034
Soda Ash, soft stage, kg/cu m = 0.243

REMARKS: Could not build up softener reaction zone solids, poor effluent clarity. No filter operation. Average operating temperature 18°C. Heavy floc carry over in sedimentation basin effluent.

AVERAGE PERFORMANCE DATA*

Parameter	Raw water	Sed. basin effluent	Softener effluent	Recarb. effluent
Flow, mgd.	3.5	3.5	3.5	3.5
Flow, cu m/day	13,248	13,248	13,248	13,248
pH units	3.0	7.7	10.2	9.9
Acid, tot.	190	--	--	--
Alk, tot.	--	16	53	54
Alk, "P"	--	0	28	23
Ca, tot.	30	73	37	37
Ca, diss.	--	69	28	25
Mg, tot.	19	22	14	12
Mg, diss.	--	20	12	12
Fe, tot.	15	2.7	.3	.3
Fe, diss.	--	0.1	.1	.1
Mn, tot.	5.0	3.4	.38	.42
Mn, diss.	--	2.0	.26	.16
Al, tot.	14	2.1	.3	.2
Al, diss.	--	0.1	.16	.13
Na	1.9	3.3	109	109
Hard. (calc.)	265	258	119	109
SO_4	273	240	233	231
Sp. C. umhos/cm^2	897	607	742	735
TSS	--	--	--	--
TDS	--	--	--	--

*Unless otherwise noted, all results are expressed in mg/l of the constituent indicated. Dissolved values were determined on samples filtered through a borosilicate microfiber glass filter. Acidity, alkalinity, and hardness are expressed as calcium carbonate equivalent values.

PERIOD: September 20 - 25, 1974

PROCESS OBJECTIVES: Neutralization Stage pH = 7.0
Softening Stage pH = 11.4
Hardness Objective = 100 mg/l

CHEMICAL FEED: Lime, neutralization stage, kg/cu m = 0.086
Lime, softening stage, kg/cu m = 0.064
Soda Ash, soft stage, kg/cu m = 0.231

REMARKS: Heavy, large floc carryover in sedimentation basin effluent. Softener reaction zone solids were generated at 11.4 pH. Floc settled well resulting in a clear softener clarification zone. Average operating temperature 15.6°C. Manganese removal was not satisfactory.

AVERAGE PERFORMANCE DATA*

Parameters	Raw water	Sed. basin effluent	Softener effluent	Recarb. effluent
Flow, mgd	3.5	3.5	3.5	3.5
Flow, cu m/day	13,248	13,248	13,248	13,248
pH, units	2.9	7.5	11.2	9.1
Acid, tot.	162	--	--	--
Alk. tot.	--	19	67	55
Alk, "P"	--	--	59	3
Ca, tot.	30	72	50	47
Ca, diss.	--	71	50	40
Mg, tot.	19	20	3	3
Mg, diss.	--	19	2.6	2.8
Fe, tot.	14	3.7	.2	.2
Fe, diss.	--	.2	.1	.1
Mn, tot.	4.6	4.5	.16	.16
Mn, diss.	--	3.6	.30	.26
Al, tot.	13	2.9	.2	.2
Al, diss.	--	.2	.3	.1
Na	2.1	2.4	103	99
Hard. (calc.)	258	260	128	112
SO_4	267	232	236	223
Sp. C. umhos/cm^2	891	620	882	700
TSS	--	--	--	31
TDS	--	--	--	--

*Unless otherwise noted, all results are expressed in mg/l of the constituent indicated. Dissolved values were determined on samples filtered through a borosilicate microfiber glass filter. Acidity, alkalinity, and hardness are expressed as calcium carbonate equivalent values.

PERIOD: September 26 - 30, 1974

PROCESS OBJECTIVES: Neutralization Stage pH = 7.0
Softening Stage pH = 10.5 - 11.0
Hardness Objective = 100 mg/l

CHEMICAL FEED: Lime, neutralization stage, kg/cu m = 0.102
Lime, softening stage, kg/cu m = 0.043
Soda Ash, soft stage, kg/cu m = 0.22

REMARKS: Softener effluent poor in clarity. Manganese too high.
Average operating temperature 15.7°C.

AVERAGE PERFORMANCE DATA*

Parameter	Raw water	Sed. basin effluent	Softener effluent	Recarb. effluent
Flow, mgd.	3.5	3.5	3.5	3.5
Flow, cu m/day	13,248	13,248	13,248	13,248
pH Units	3.1	6.7	10.6	8.0
Acid, tot.	176	--	--	--
Alk, tot.	--	25	41	60
Alk, "P"	--	--	--	--
Ca, tot.	30	73	31	28
Ca, diss.	--	73	29	30
Mg, tot.	20	23	15	10
Mg, diss.	--	22	14	10
Fe, tot.	19	3.7	0.5	.6
Fe, diss.	--	0.1	0.1	.1
Mn, tot.	5.0	5.2	.23	.3
Mn, diss.	--	4.2	.28	.4
Al, tot.	14	2.9	.4	.5
Al, diss.	--	0.2	.3	.3
Na	2.1	2.4	98	99
Hard. (calc.)	277	278	112	110
SO_4	289	235	240	229
Sp. C umhos/cm^2	974	640	729	735
TSS	--	--	--	--
TDS	--	--	--	--

*Unless otherwise noted, all results are expressed in mg/l of the constituent indicated. Dissolved values were determined on samples filtered through a borosilicate microfiber glass filter. Acidity, alkalinity, and hardness are expressed as calcium carbonate equivalent values.

PERIOD: October 1 - 4, 1974

PROCESS OBJECTIVES: Neutralization Stage pH = 7.0
Softening Stage pH = 11.4
Hardness objective = 100 mg/l
Controlling level in aeration tank and regulating aerator operation (2 hrs/shift) to reduce floc breakup and improve solids removal in sedmintation basin.

CHEMICAL FEED: Lime, neutralization stage, kg/cu m = 0.102
Lime, softening stage, kg/cu m = 0.061
Soda Ash, soft stage, kg/cu m = 0.234

REMARKS: Excellent softener operation, 14% by volume settleable reaction zone solids (5 min. test). No significant improvement in sed. basin solids removal. Average operating temperature 12.1°C in softener.

AVERAGE PERFORMANCE DATA*

Parameter	Raw water	Sed. basin effluent	Recarb. effluent
Flow, mgd	3.5	3.5	3.5
Flow, cu m/day	13,248	13,248	13,248
pH units	2.9	7.0	11.4
Acid, tot.	178	--	--
Alk, tot.	--	25	58
Alk, "P"	--	--	--
Ca, tot.	31	78	44
Ca, diss.	--	77	41
Mg, tot.	19	21	2.0
Mg, diss.	--	19	2.8
Fe, tot.	15	4.0	0.1
Fe, diss.	--	0.1	0.1
Mn, tot.	4.8	4.9	.08
Mn, diss.	--	4.0	--
Al, tot.	13	3.0	0.1
Al, diss.	--	0.1	0.1
Na	2.2	2.5	104
Hard. (calc.)	267	277	112
SO_4	278	231	228
Sp. C. umhos/cm^2	948	610	840
TSS	--	--	4
TDS	--	--	--

*Unless otherwise noted, all results are expressed in mg/l of the constituent indicated. Dissolved values were determined on samples filtered through a borosilicate microfiber glass filter. Acidity, alkalinity, and hardness are expressed as calcium carbonate equivalent values.

PERIOD: October 5 - 7, 1974

PROCESS OBJECTIVE: Neutralization Stage pH = 7.0
Softening Stage pH = 11.5
Hardness Objective = 100 mg/l
Coagulant aid addition (1 mg/l) to sed. basin to improve solids removal

CHEMICAL FEED: Lime, Neutralization stage, kg/cu m = 0.113
Lime, softening stage, kg/cu m = 0.053
Soda Ash, soft stage, kg/cu m = 0.217
Coag. Aid, sed. basin, kg/cu m = 0.001

REMARKS: Excellent softener operation, 13% (by volume) settleable reaction zone solids. Coagulant aid improved sed. basin solids removal. Average operating temperature 12.5°C.

AVERAGE PERFORMANCE DATA*

Parameter	Raw water	Sed. basin effluent	Recarb. effluent
Flow, mgd.	3.5	3.5	3.5
Flow, cu m/day	13,248	13,248	13,248
pH units	2.9	7.0	11.4
Acid, tot.	186	--	--
Alk, tot.	--	14	84
Alk, "P"	--	--	80
Ca, tot.	32	78	52
Ca, diss.	--	73	48
Mg, tot.	20	20	1.8
Mg, diss.	--	18	2.1
Fe, tot.	17	1.6	0.1
Fe, diss.	--	0.1	0.1
Mn, tot.	5.0	3.7	.04
Mn, diss.	--	3.3	--
Al, tot.	13	1.7	0.1
Al, diss.	--	0.1	0.1
Na	2.0	2.5	97
Hard. (calc.)	276	272	127
SO_4	293	230	225
Sp. C. umhos/cm^2	963	595	990
TSS	--	8	3
TDS	--	--	--

*Unless otherwise noted, all results are expressed in mg/l of the constituent indicated. Dissolved values were determined on samples filtered through a borosilicate microfiber glass filter. Acidity, alkalinity, and hardness are expressed as calcium carbonate equivalent values.

PERIOD: October 8 - 10, 1974

PROCESS OBJECTIVES: Neutralization Stage pH = 7.0
Softening Stage pH = 11.3
Hardness Objective = 100 mg/l
Feeding 6.3 mg/l potassium permanganate to softener for manganese oxidation.

CHEMICAL FEED: Lime, Neutralization stage, kg/cu m = 0.108
Lime, softening stage, kg/cu m = 0.053
Soda Ash, soft stage, kg/cu m = 0.204
Potass. Perm., soft stage, kg/cu m = 0.006

REMARKS: Good softener operation, but unable to reduce manganese to acceptable level. Average operating temperature 12.5°C. 12% by volume settleable reaction zone solids.

AVERAGE PERFORMANCE DATA*

Parameter	Raw water	Sed. basin effluent	Recarb. effluent
Flow, mgd.	3.5	3.5	3.5
Flow, cu m/day	13,248	13,248	13,248
pH units	3.0	7.3	11.3
Acid, tot.	179	--	--
Alk, tot.	--	14	54
Alk, "P"	--	--	41
Ca, tot.	31	72	51
Ca, diss.	--	70	47
Mg, tot.	20	19	2.0
Mg, diss.	--	18	3.1
Fe, tot.	15	3.0	0.1
Fe, diss.	--	0.2	0.1
Mn, tot.	48	4.2	0.27
Mn, diss.	--	3.9	0.27
Al, tot.	14	2.3	0.2
Al, diss.	--	0.1	0.2
Na	2.3	2.5	88
Hard. (calc)	268	257	123
SO_4	277	212	199
Sp. C. umhos/cm^2	902	550	773
TSS	--	12	5
TDS	--	--	--

*Unless otherwise noted, all results are expressed in mg/l of the constituent indicated. Dissolved values were determined on samples filtered through a borosilicate microfiber glass filter. Acidity, alkalinity, and hardness are expressed as calcium carbonate equivalent values.

PERIOD: October 12 - 16, 1974

PROCESS OBJECTIVES: Neutralization Stage pH = 9.5
Softening Stage pH = 11.3
Hardness Objective = 100 mg/l

CHEMICAL FEED: Lime, neutralization stage, kg/cu m = 0.119
Lime, softening stage, kg/cu m = 0.023
Soda Ash, soft stage, kg/cu m = 0.222

REMARKS: Poor softener operation, no reaction zone solids. Average operating temperature 13.6°C. (Note: lower magnesium and manganese levels in sed. basin effluent)

AVERAGE PERFORMANCE DATA*

Parameter	Raw water	Sed. basin effluent	Recarb. effluent
Flow, mgd.	3.5	3.5	3.5
Flow, cu m/day	13,248	13,248	13,248
pH units	3.1	9.7	11.3
Acid, tot.	160	--	--
Alk, tot.	--	18	67
Alk, "P"	--	--	46
Ca, tot.	31	89	60
Ca, diss.	--	89	45
Mg, tot.	17	7.5	2.6
Mg, diss.	--	6.8	2.0
Fe, tot.	13	1.5	0.1
Fe, diss.	--	0.1	0.1
Mn, tot.	4.5	.29	.09
Mn, diss.	--	.08	.01
Al, tot.	13	2.9	1.8
Al, diss.	--	1.2	1.0
Na	2.1	2.4	99
Hard. (calc.)	244	259	126
SO_4	247	198	198
Sp. C umhos/cm^2	794	546	799
TSS	--	--	49
TDS	--	--	--

*Unless otherwise noted, all results are expressed in mg/l of the constituent indicated. Dissolved values were determined on samples filtered through a borosilicate microfiber glass filter. Acidity, alkalinity, and hardness are expressed as calcium carbonate equivalent values.

PERIOD: October 17 - 22, 1974

PROCESS OBJECTIVES: Neutralization Stage pH = 9.5
Softening Stage pH = 11.4
Hardness Objective = 200 mg/l

CHEMICAL FEED: Lime, neutralization stage, kg/cu m = 0.087
Lime, softening stage, kg/cu m = 0.027
Soda Ash, soft stage, kg/cu m = 0.137
Coag. Aid, soft stage, kg.cu m = 0.001

REMARKS: No reaction zone solids build up in softener, effluent clarity poor. Average operating temperature 10°C. No improvement with the use of coagulant aid.

AVERAGE PERFORMANCE DATA*

Parameter	Raw water	Sed. basin effluent	Recarb. effluent
Flow, mgd.	3.5	3.5	3.5
Flow, cu m/day	13,248	13,248	13,248
pH units	3.1	9.3	11.1
Acid, tot.	160	--	--
Alk, tot.	--	15	70
Alk, "P"	--	--	56
Ca, tot.	30	82	90
Ca, diss.	--	82	76
Mg, tot.	17	14	3.9
Mg, diss.	--	11.3	3.7
Fe, tot.	11.6	2.2	0.1
Fe, diss.	--	0.1	0.1
Mn, tot.	4.2	.69	.04
Mn, diss.	--	.10	.02
Al, tot.	10.5	2.8	0.2
Al, diss.	--	1.0	0.2
Na	2.4	2.6	62
Hard. (calc.)	226	256	207
SO_4	224	206	207
Sp. C. umhos/cm^2	746	555	834
TSS	--	--	39
TDS	--	--	--

*Unless otherwise noted, all results are expressed in mg/l of the constituent indicated. Dissolved values were determined on samples filtered through a borosilicate microfiber glass filter. Acidity, alkalinity, and hardness are expressed as calcium carbonate equivalent values.

PERIOD: October 28 - November 11, 1974

PROCESS OBJECTIVES: Neutralization Stage pH = 8.5 - 9.0
Softening Stage pH = 11.3
Hardness Objective = 200 mg/l

CHEMICAL FEED: Lime, neutralization stage, kg/cu m = 0.072
Lime, softening stage, kg/cu m = 0.036
Soda Ash, soft stage, kg/cu m = 0.112

REMARKS: Average operating temperature (softener) 11°C. Flow to filters limited to 2.0 mgd applied to one-fourth of the filter capacity. In this way, performance could be evaluated at the hydraulic design loading.

AVERAGE PERFORMANCE DATA*

Parameters	Raw water	Sed. basin effluent	Mix Box	Softener effluent	Filter effluent
Flow, mgd	3.5	3.5	3.5	3.5	2.0
Flow, cu m/day	13,248	13,248	13,248	13,248	7,570
pH, units	3.0	7.2	10.7	10.7	7.9
Acid, tot.	143	--	--	--	--
Alk, tot.	--	15	90	39	53
Alk, "P"	--	--	35	33	--
Ca, tot.	31	76	112	73	65
Ca, diss.	--	75	--	65	--
Mg, tot.	16	15	14	4.3	3.8
Mg, diss.	--	15	--	4.3	--
Fe, tot.	13	2.1	1.9	0.1	0.1
Fe, diss.	--	0.1	--	0.1	--
Mn, tot.	4.3	1.4	1.19	.08	.03
Mn, diss.	--	0.8	--	.08	--
Al, tot.	10.4	2.2	1.9	0.1	0.1
Al, diss.	--	0.6	--	.07	--
Na	2.4	2.6	2.8	51	46
Hard. (calc.)	228	253	357	182	176
SO_4	226	215	211	211	197
Sp. C umhos/cm^2	774	572	630	670	623
TSS	--	--	--	8	--
TDS	--	--	--	--	--

*Unless otherwise noted, all results are expressed in mg/l of the constituent indicated. Dissolved values were determined on samples filtered through a borosilicate microfiber glass filter. Acidity, alkalinity, and hardness are expressed as calcium carbonate equivalent values.

PERIOD: November 11 - 17, 1974 (6 days, 11/21/75 omitted, see remarks below)

PROCESS OBJECTIVES: Neutralization Stage pH = 8.5 - 9.0
Softening Stage pH = 11.3
Hardness Objective = 200 mg/l
Demanganization by feeding potassium permanganate (10 mg/l) to the influent raw water line.

CHEMICAL FEED: Lime, neutralization stage, kg/cu m = 0.088
Lime, softening stage, kg/cu m = 0.026
Soda Ash, softening stage, kg/cu m = 0.084
Potass. perm., influ. line, kg/cu m = 0.041

REMARKS: Heavy rainfall occurred on November 12 causing a 5.0 mgd flow through the plant for a short period. This hydraulic overload upset the softener operation. Average softener operating temperature 6.4°C. Filter runs were 54 hours.

AVERAGE PERFORMANCE DATA*

Parameter	Raw water	Sed. Basin effluent	Mix Box	Softener effluent	Recarb. effluent	Filter effluent
Flow, mgd	3.5	3.5	3.5	3.5	3.5	2.0
Flow, cu m /day	13,248	13,248	13,248	13,248	13,248	7,570
pH, units	3.1	7.0	10.5	10.5	8.4	8.1
Acid, tot.	103	--	--	--	--	--
Alk, tot.	--	14	80	35	49	46
Alk, "P"	--	--	20	27	--	--
Ca, tot.	28	80	111	80	74	74
Ca, diss.	--	80	--	77	--	--
Mg, tot.	14	16	16	6.6	6.8	6.6
Mg, diss.	--	15	--	6.6	--	--
Fe, tot.	6	1.8	1.8	0.2	0.2	0.1
Fe, diss.	--	0.2	--	0.1	--	--
Mn, tot.	3.4	.74	.67	.08	.06	.02
Mn, diss.	--	.16	--	.03	--	--
Al, tot.	8.7	1.9	1.9	0.2	0.2	0.1
Al, diss.	--	0.5	--	0.1	--	--
Na	3.3	3.2	3.2	39	41	40
Hard. (Calc.)	194	264	356	221	216	214
SO_4	183	233	233	234	225	225
Sp. C umhos/cm^2	638	627	702	714	691	683
TSS	--	--	--	42	--	--
TDS	--	--	--	--	--	--

*Unless otherwise noted, all results are expressed in mg/l of the constituent indicated. Dissolved values were determined on samples filtered through a borosilicate microfiber glass filter. Acidity, alkalinity, and hardness are expressed as calcium carbonate equivalent values.

PERIOD: November 18 - 21, 1974 (4 days)

PROCESS OBJECTIVES: Neutralization Stage pH = 8.5 - 9.0
Softening Stage pH = 10.5
Hardness Objective = 200 mg/l

CHEMICAL FEED: Lime, neutralization stage, kg/cu m = 0.11
Lime, softening stage, kg/cu m = 0.016
Soda Ash, soft stage, kg/cu m = 0.10
Potass. perm., influ. line kg/cu m = 0.01

REMARKS: Softener clarification poor with brown floc carry over.
Average operating temperature 5.3°C. Filter runs 30 hours.

AVERAGE PERFORMANCE DATA*

Parameter	Raw water	Sed. Basin effluent	Mix box	Softener effluent	Recarb. effluent	Filter effluent
Flow, mgd	3.5	3.5	3.5	3.5	3.5	2.0
Flow, cu m/day	13,248	13,248	13,248	13,248	13,248	7,570
pH, units	2.8	7.8	9.4	9.2	7.7	6.9
Acid, tot.	193	--	--	--	--	--
Alk, tot.	--	14	46	24	41	43
Alk, "P"	--	--	3	2	--	--
Ca, tot.	30	92	96	70	64	63
Ca, diss.	--	85	--	64	--	--
Mg, tot.	18	16	17	16	13	13
Mg, diss.	--	16	--	13	--	--
Fe, tot.	15	3.4	2.9	1.8	1.2	0.3
Fe, diss.	--	0.1	--	0.1	--	--
Mn, tot.	4.6	.98	.86	.48	0.4	.04
Mn, diss.	--	.04	--	.03	--	--
Al,tot.	15	2.6	2.3	1.2	0.8	0.1
Al, diss.	--	0.6	--	0.3	--	--
Na	2.7	3.1	4.7	46	46	46
Hard. (calc.)	270	284	329	218	223	212
SO_4	275	245	247	244	229	240
Sp. C. umhos/cm^2	919	610	616	645	680	675
TSS	--	--	--	--	26	--
TDS	--	--	--	--	--	460

*Unless otherwise noted, all results are expressed in mg/l of the constituent indicated. Dissolved values were determined on samples filtered through a borosilicate microfiber glass filter. Acidity, alkalinity, and hardness are expressed as calcium carbonate equivalent values.

PERIOD: November 22 - 24, 1974 (3 days)

PROCESS OBJECTIVES: Neutralization Stage pH = 8.5 - 9.0
Softening Stage, pH = 10.5
Hardness Objective = 200 mg/l

CHEMICAL FEED: Lime, neutralization stage, kg/cu m = 0.104
Lime softening stage, kg/cu m = 0.018
Soda Ash, soft stage, kg/cu m = 0.166
Carbon dioxide, recarb., kg/cu m = 0.044

REMARKS: Softener clarification poor. Average operating temperature 5°C
Filter runs 20 hours.

AVERAGE PERFORMANCE DATA*

Parameter	Raw water	Sed. basin effluent	Mix box	Softener effluent	Recarb. effluent	Filter effluent
Flow, mgd	3.5	3.5	3.5	3.5	3.5	2.0
Flow, cu m/day	13,248	13,248	13,248	13,248	13,248	7,570
pH, units	3.0	7.0	9.5	9.7	8.3	7.9
Acid, tot.	175	--	--	--	--	--
Alk, tot.	--	14	38	103	105	84
Alk, "P"	--	--	3	9	0	--
Ca, tot.	29	94	88	67	59	45
Ca, diss.	--	86	--	43	--	--
Mg, tot.	18	16	18	13	12	12
Mg, diss.	--	16	--	12	--	--
Fe, tot.	18	3.5	3.0	1.5	1.3	0.2
Fe, diss.	--	0.1	--	0.1	--	--
Mn, tot.	4.4	1.5	1.2	.46	.52	.04
Mn, diss.	--	.36	--	.04	--	--
Al, tot.	15	2.7	2.5	1.2	1.0	0.1
Al, diss.	--	0.4	--	0.4	--	--
Na	2.9	3.6	--	75	73	73
Hard. (calc.)	269	287	315	159	--	165
SO_4	260	244	245	239	233	231
Sp. C. umhos/cm^2	893	613	617	697	767	757
TSS	--	--	--	--	46	--
TDS	--	--	--	--	--	747

*Unless otherwise noted, all results are expressed in mg/l of the constituent indicated. Dissolved values were determined on samples filtered through a borosilicate microfiber glass filter. Acidity, alkalinity, and hardness are expressed as calcium carbonate equivalent values.

PERIOD: November 25 - 30, 1974 (6 days)

PROCESS OBJECTIVES: Neutralization Stage pH = 8.5 - 9.0
Softening Stage pH = 10.5
Hardness Objective = 106 mg/l

CHEMICAL FEED: Lime, Neutralization stage, kg/cu m = 0.098
Lime, softening stage, kg/cu m = 0.037
Soda Ash, soft stage, kg/cu m = 0.33
Carbon dioxide, recarb., kg/cu m = 0.075 (11/25 & 11/26 only)

REMARKS: Poor softener clarification - 20 hour filter runs. Average operating temperature 2.3°C.

AVERAGE PERFORMANCE DATA*

Parameter	Raw water	Sed. basin effluent	Mix box	Softener effluent	Recarb. effluent	Filter effluent
Flow, mgd	3.5	3.5	3.5	3.5	3.5	2.0
Flow, cu m/day	13,248	13,248	13,248	13,248	13,248	7,470
pH, units	2.9	7.4	9.4	10.4	8.2	8.4
Acid, tot.	173	--	--	--	--	--
Alk, tot.	--	24	91	153	177	160
Alk, "P"	--	--	6	43	0	2
Ca, tot.	27	94	104	44	38	25
Ca, diss.	--	88	--	28	--	--
Mg, tot.	17	18	18	9.9	9.8	9.3
Mg, diss.	--	18	--	9.4	--	--
Fe, tot.	24	3.2	2.9	0.7	0.7	0.1
Fe, diss.	--	0.1	--	0.1	--	--
Mn, tot.	4.2	2.2	2.0	.44	.34	.08
Mn, diss.	--	1.2	--	.08	--	--
Al, tot.	15	2.7	2.4	0.5	0.4	0.1
Al, diss.	--	0.5	--	0.2	--	--
Na	2.3	2.8	2.4	146	137	141
Hard. (calc.)	272	297	359	112	139	103
SO_4	262	257	254	254	234	235
Sp. C umhos/cm^2	890	646	734	941	917	867
TSS	--	--	--	--	53	--
TDS	--	--	--	--	--	--

*Unless otherwise noted, all results are expressed in mg/l of the constituent indicated. Dissolved values were determined on samples filtered through a borosilicate microfiber glass filter. Acidity, alkalinity, and hardness are expressed as calcium carbonate equivalent values.

PERIOD: December 3 - 8, 1974 (6 days)

PROCESS OBJECTIVES: Neutralization Stage pH = 7.0 - 7.5
Softening Stage pH = 11.3
Hardness Objective = 100 mg/l

CHEMICAL FEED: Lime, Neutralization stage, kg/cu m = 0.098
Lime, softening stage, kg/cu m = 0.053
Soda Ash, soft stage, kg/cu m = 0.317
Carbon dioxide, recarb., kg/cu m = 0.18 (12/5 - 12/8)

REMARKS: Poor softener clarification. Filter runs 20 hours. Average operating temperature 1.5°C.

AVERAGE PERFORMANCE DATA*

Parameter	Raw water	Sed. basin effluent	Mix box	Softener effluent	Recarb. effluent	Filter effluent
Flow, mgd	3.5	3.5	3.5	3.5	3.5	2.0
Flow cu m/day	13,248	13,248	13,248	13,248	13,248	7,570
pH Units	2.9	7.1	10.2	10.6	7.8	7.6
Acid, tot.	190	--	--	--	--	--
Alk., tot.	--	24	106	175	96	96
Alk., "P"	--	--	18	34	0	--
Ca, totl.	26	96	130	64	62	54
Ca, diss.	--	90	--	40	--	--
Mg, tot.	16	19	16	7.6	6.7	7.3
Mg, diss.	--	18	--	5.6	--	--
Fe, tot.	26	3.7	2.4	1.1	0.5	0.1
Fe, diss.	--	0.2	--	0.1	--	--
Mn, tot.	4.0	7.7	7.2	.98	.33	.19
Mn, diss.	--	4.8	--	.37	--	--
Al, tot.	16	2.4	1.9	0.2	0.2	0.1
Al, diss.	--	0.1	--	0.1	--	--
Na	2.0	2.0	2.7	140	111	116
Hard.(calc.)	268	312	426	133	--	--
SO_4	268	274	270	271	258	292
Sp.C umhos/cm^2	915	682	755	945	853	922
TSS	--	--	--	--	41	--
TDS	--	--	--	--	--	--

*Unless otherwise noted, all results are expressed in mg/l of the constituent indicated. Dissolved values were determined on samples filtered through a borosilicate microfiber glass filter. Acidity, alkalinity, and hardness are expressed as calcium carbonate equivalent values.

PERIOD: June 16-26, 1975 (10 days)

PROCESS OBJECTIVES:
Neutralization Stage pH = 9.0
Softening Stage pH = 9.6
Hardness Objective = 100 mg/l
Recarbonation effluent pH = 8.5

CHEMICAL FEED:
Lime Neutralization stage, kg/cu m = 0.138
Lime, softening stage, kg/cu m = 0
Soda Ash, soft. stage, kg/cu m = 0.334
Coag. Aid., Sed. Basin, kg/cu m = 0.0004
Carbon Dioxide, recarb., kg/cu m = 0.029

REMARKS: Temperature ranged in softener from 12°-20"C (Avg. 17°C) Clarification zone was hazy/turbid with finely divided precipitate in suspension. Hardness objective (100 mg/l) was not met.

AVERAGE PERFORMANCE DATA*

Parameter	Raw water	Sed. basin effluent	Softener effluent	Recarb. effluent
Flow, mgd	3.5	3.5	3.5	3.5
Flow cu m/day	13,248	13,248	13,248	13,248
pH, units	2.1	7.5	9.4	8.7
Acid, tot.	191	--	--	--
Alk., tot.	--	-1	53	82
Alk, "P"	--	10	9	3
Ca, tot.	28	113	40	39
Ca, diss.	--	111	32	31
Mg, tot.	20	16	12	12
Mg, diss.	--	16	11	11
Fe, tot.	23	1.4	.29	0.3
Fe, diss.	--	.07	.03	.06
Mn, tot.	4.9	1.8	.43	.32
Mn, diss.	--	0.9	.04	.07
Al, tot.	17	2.5	1.2	1.2
Al, diss.	--	1.5	.7	0.5
Na	2.9	1.6	138	136
Hard.(calc.)	252	364	127	124
SO_4	339	322	315	305
Sp. C umhos/cm^2	807	618	814	825
TSS	--	--	--	--
TDS	--	--	--	573

*Unless otherwise noted, all results are expressed in mg/l of the constituent indicated. Dissolved values were determined on samples filtered through a borosilicate microfiber glass filter. Acidity, alkalinity, and hardness are expressed as calcium carbonate equivalent values.